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Abstract

CSIRO's patented CS-Cap process aims at reducing the costs of amine-based post-combustion capture by combining SO2 and CO2 capture using one absorbent in a single absorber column. By avoiding the need for a separate flue gas desulfurization unit, the process offers potential savings for power plants requiring CO2 capture. High-level cost estimates based on lab and pilot data are presented for two amine reclamation techniques i.e. thermal reclamation and reactive crystallisation. Only regeneration via reactive crystallisation reduces CS-Cap costs below base case FGD/SCR-PCC. Cost estimations suggest a potential reduction of 38–44% in the total plant cost when using the CS-Cap process compared to base case. However, the amine reclaimer operating cost governs the overall cost of the CS-Cap process and is highly sensitive to sulfur content. A 50% reduction is observed when SO2 levels reduce from 700 to 200 ppm. Comparing levelised cost of electricity and CO2 avoided costs for CS-Cap against our base case, low sulfur brown coal has a slight (5-7%) cost advantage; however, confirmation requires pilot data on amine recovery.

Keywords	thermal reclamation; reactive crystallization; total plant cost; Aspen Plus Economic Analyzer; cost estimation
Taxonomy	Computational Chemistry, Applied Sciences, Chemical Engineering
Corresponding Author	Vincent Verheyen
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Suggested reviewers	Jon Gibbins, Fernando Vega, Earl Goetheer

Submission Files Included in this PDF

File Name [File Type]

Bharti et al IGGC Cover Letter - costing revised manuscript 5-5-20.docx [Cover Letter]

Costing paper final May 2020 reveal changes.docx [Response to Reviewers]

highlights.docx [Highlights]

Costing paper final May 2020.docx [Manuscript File]

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Research Data Related to this Submission

There are no linked research data sets for this submission. The following reason is given:
The data that has been used is confidential

5th May 2020

Dr. Ramirez

Associate Editor

International Journal of Greenhouse Gas Control

RE: Revision requested for JGGC_2019_643_R1

Please find the revised manuscript enclosed along with our responses to the revision requested for JGGC_2019_643:

Title: Techno-economic evaluation of amine-reclamation technologies and combined CO2/SO2 capture for Australian coal-fired plants

Corresponding author: T. Vincent Verheyen

Co-authors: Bharti Garg, Nawshad Haque, Ashleigh Cousins, Pauline Pearson, Paul HM Feron

We have carefully considered all issues mentioned in the reviewers' comments, and outline every change made point by point, and provide suitable rebuttals for any comments below: We believe it has further benefited from the changes suggested by the additional reviewer and ask you to reconsider its publication in the International Journal of Greenhouse Gas Control.

I have included below our response to the reviewers comments. In addition, a “track changes” version of the revised manuscript is included in the pdf build along with a “clean” version.

Comments from the reviewer:

-Reviewer 2

- Thank to the editor for giving me the opportunity to review this paper. In general, the paper is a well-English written manuscript and also the results and conclusions are highly valuable for CO2 capture develpment for PC power plants.

I recomend its publication after minor revision of the final version of the paper. The main comment to the authors are listed below:

1) The thermal reclaimer (option 1) occurs at elevated temperatures (around 170ºC). At this temperature level, MEA degrades based on thermal degradation mechanism. Were thermal degradation of MEA considered in the techno-economic analysis of option 1? Please, discuss about this in the manuscript and also add some important references about degradation compounds of MEA (i.e. works from Goudegard, Lepaumier, Vega, etc)

Agreed Option 1 text on page 7 has been changed to include discussion on thermal degradation and references suggested by the reviewer. The temperatures used in our research are much lower than 170C and oxidative degradation is expected to be the dominant cause of MEA loss

2) NO is mentioned in the result section but a more detailed description of the NO treatment n Australian PCC´s should be discussed. NO also degrades MEA (please see references from Prof Maroto-Valer review in 2014). It should be also taken into account.

Agreed Assuming reviewer meant to say NOx. Extra discussion (page 20) on NO as the major NOx component in Australian flue gases and additional reference added

Minor comments:

1) Not use capital letter for electrodialysis in the final paragraph of section 1.1 agreed see P.4

2) SIFigure is not cited adequately in the introduction section disagree as the authors have created a new figure for this paper hence not cited, the process is presented diagrammatically in other cited papers but; is we feel more clearly presented here in our own figure

3) Why did not the authors consider oxadolizone as a main degradation product of MEA instead of formate? Please, discuss in the manuscript and add references yes now discussed and referenced see P. 9-10

4) Please, correct units in Table 2 for operating costs accepted see P.17

5) Table 5 appeared in the manuscript before Table 4. Please, correct it Accepted & corrected by replacing with Fig.2 as in point 6 below

6) Table 5 should be transformed to a Figure for easier comprehension agreed see P. 24 now has additional Fig. 2

7) The last paragraph in pag 17 (about the strategy for scale-up the costs of equipment) should be moved to the cost methodology section accepted section moved to P. 15

If you require any further information, Vincent can be contacted by email (Vince.Verheyen@federation.edu.au) or telephone +61 42 322 595.

We thank you for your time and consideration.

Best regards,

Vincent Verheyen, on behalf of all the authors of this publication.

Techno-economic evaluation of amine-reclamation technologies and combined CO_2/SO_2 capture for Australian coal-fired plants

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Abstract

CSIRO's patented CS-Cap process aims at reducing the costs of amine-based post-combustion capture by combining SO_2 and CO_2 capture using one absorbent in a single absorber column. By avoiding the need for a separate flue gas desulfurization unit, the process offers potential savings for power plants requiring CO_2 capture. High-level cost estimates based on lab and pilot data are presented for two amine reclamation techniques i.e. thermal reclamation and reactive crystallisation. Only regeneration via reactive crystallisation reduces CS-Cap costs below base case FGD/SCR-PCC. Cost estimations suggest a potential reduction of 38–44% in the total plant cost when using the CS-Cap process compared to base case. However, the amine reclaimer operating cost governs the overall cost of the CS-Cap process and is highly sensitive to sulfur content. A 50% reduction is observed when SO_2 levels reduce from 700 to 200 ppm. Comparing levelised cost of electricity and CO_2 avoided costs for CS-Cap against our base case, low sulfur brown coal has a slight (5-7%) cost advantage; however, confirmation requires pilot data on amine recovery.

Keywords: CS-Cap; thermal reclamation; reactive crystallisation; total plant cost; Aspen Plus Economic Analyser; cost estimation

1. Introduction

Approximately 70% of Australia's electricity generation comes from local black and brown coal (Engineers Australia, 2017). However, coal-based electricity generation is facing significant emission constraints due to the need for climate change mitigation.

Future energy generation and security will not be solely dependent on any one power technology. Instead, a strategic approach that combines existing generation with emerging low-emission technologies will lead to a smooth and sustainable transition. As per the Australian Government's Energy White Paper (Engineers Australia, 2017), coal could continue supplying energy, provided emissions are reduced at low cost. Integrating carbon capture and storage (CCS) into power stations will significantly reduce emissions, but will result in high-cost electricity. In the net zero CO_2 by 2050 scenario, CCS needs fast deployment and that requires lower-cost capture technologies. This paper investigates such a technology i.e. a single column SO_2 and CO_2 capture process, and estimates the cost of key process and the amine regeneration scenarios.

1.1. CSIRO's CS-Cap process

Amine-based Post Combustion Capture (PCC) is currently the leading technology for capturing CO_2 from power plants (IEAGHG, 2019). Pre-treatment of flue gases, including removal of SO_2, is essential for efficient PCC. SO_2 preferentially reacts with amines to form degradation products, thus increasing amine makeup requirements (CO2CRC, 2017). The prior removal of SO_2, requires flue gas desulfurisation (FGD). In Northern Hemisphere plants, FGD is typically conducted via a wet limestone process (Jamil et.al, 2013). Amine based PCC requires FGD units to reduce SO_2 levels below 10 ppm to minimise degradation. However, existing FGD units may not reduce SO_2 levels to such low levels, and these plants require additional scrubbers or deep FGD units (Puxty et al., 2014). Australian plants do not use FGD units (Puyvelde, 2009), therefore, any PCC retrofit requires additional investment (Rubin, Davison, & Herzog, 2015).

As an alternative to high-cost FGD or upgraded FGD, CSIRO has developed a combined SO_2 + CO_2 capture process utilising a single aqueous amine solution in a single absorber column (Beyad et al., 2014). Known as CS-Cap, the process offers potential capex savings by avoiding the need for a separate FGD unit (Puxty et al., 2014). A proof of concept operation was conducted at CSIRO's pilot facility, at AGL's Loy Yang brown-coal-fired power station (Pearson et al., 2017). A flow diagram of the CS-Cap process, which has been patented by CSIRO, is provided in SI Figure A, the absorber captures CO_2 and SO_2 in upper and lower sections of the absorber column, respectively. The selectivity of amines to absorb stronger acids allows SO_2 to be absorbed in the bleed stream despite the presence of CO_2. Only a small percentage of the total volumetric flowrate (~0.01–3%; Beyad et al., 2014) of the CO_2-rich stream (red stream) is required to capture the incoming SO_2 in the flue gas. The CO_2-rich amine from the top section of the absorber column is regenerated by a standard steam stripping process. However, it is not possible to thermally regenerate the SO_2-rich amine formed in the lower section of the absorber as it forms a heat-stable sulfate (Pearson et al., 2017). Strong ionic bonding between protonated amines and sulfate ensures these salts do not revert to amine and SO_2 under the optimal heat treatment conditions for CO_2 release. Hence, a part of the recycle stream exiting from the bottom section of the absorber column is sent to a separate regeneration section, which is the sulfur-rich amine regeneration unit (Figure A).

The composition of the sulfur-rich CS-Cap stream sent for amine regeneration is different from a typical reclaimer stream, given its much higher (11–12 wt% versus 2 wt%) heat-stable salt burden (Garg et al., 2018). Here, the regeneration of amine can occur through thermal reclamation (ElMoudir, Supap, & Saiwan, 2012) but also through non-conventional regeneration methods.

Alternative techniques such as electrodialysis, ion exchange, nanofiltration and crystallisation are capable of regenerating amines (Garg et al., 2018) from their sulfate salts.

1.2. Investigating the cost effectiveness of the CS-Cap process

Adding CCS to a coal-fired power plant doubles the total plant cost (TPC) regardless of coal type (APGT, 2015). The variable operation and maintenance (O&M) cost is expected to rise even higher (around 3–4 times) if CCS is integrated into a new power plant. This increases the levelised cost of electricity (LCOE) from a pulverised supercritical or ultra-supercritical power plant by approximately 100% (APGT, 2015). This cost includes FGD unit installations, which as mentioned above, are a pre-requisite for any amine-based PCC technology deployed in Australia.

The huge increase in the capital and O&M costs of a new power plant built with CCS increases electricity prices while decreasing plant efficiency. This poses a serious barrier to CCS commercialisation in Australia. As described above, CSIRO's CS-Cap technology aims at reducing the capital cost of amine-based PCC by eliminating the capital cost of a separate FGD unit. However, the cost effectiveness of regenerating the CO_2 and SO_2-rich amine still needs to be investigated and compared with a standard FGD+PCC system.

As shown in Figure A (SI), amine regeneration is chiefly required for two streams in the CS-Cap process: the CO_2-rich amine stream (shown in red) and the SO_2 rich amine stream (shown in green). The CO_2-rich stream is regenerated in a standard stripper using steam. The cost of regenerating the sulfur-rich amine stream has scope for further reduction to that afforded by thermal reclamation. This is due to its unique nature (highly concentrated with sufur in form of sulfate), making it applicable to alternative lower energy reclamation techniques (Garg et al., 2018b).

The practicality and cost effectiveness of thermal reclamation and crystallisation based regeneration technologies are experimentally evaluated here using a sulfur-rich CS-Cap absorbent. This paper includes a cost estimation of these regeneration techniques, using an Aspen Plus simulation, validated against the results of our lab experiments. The cost of these regeneration techniques is integrated individually with the unit operations shown in Figure A (SI) to evaluate the overall process. The cost of CS-Cap is compared with the cost of a power plant retrofitted with:

- FGD + PCC
- FGD + PCC + improved solvent
- Minimal FGD + PCC.

2. Cost evaluation method: CS-Cap regeneration techniques

2.1. Process description and model development

Figure 1 represents the CS-Cap process with the two selected options for sulfur-rich amine regeneration. Options 1 and 2 show the thermal reclamation and reactive crystallisation units, respectively. Each option was separately integrated with the combined absorption section (SO_2 and CO_2) and CO_2-regeneration section to determine the overall cost of the CS-Cap process.

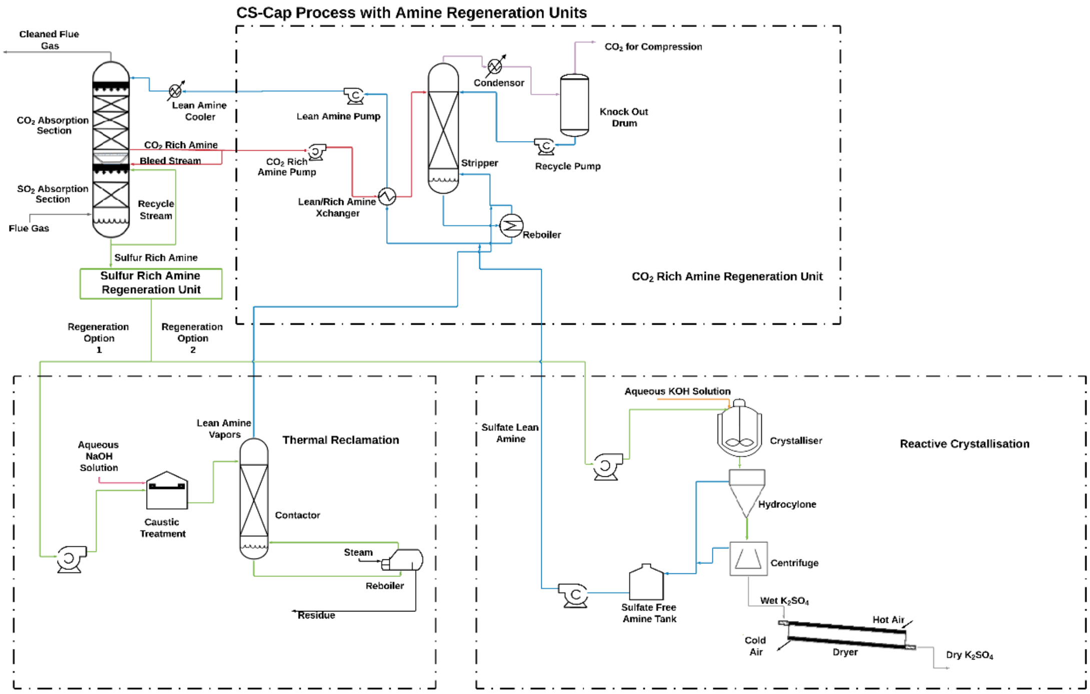


Figure 1: CS-Cap process showing reclamation of sulfur-rich amine through thermal reclamation or crystallisation

2.1.1. Option 1: Thermal reclamation

Compared to conventional HSS loadings, the more concentrated CS-Cap stream could require less or a similar amount of energy to recover the amine (Sexton et al., 2014, Wang, Hovland, & Jens, 2014). This makes the thermal reclaiming technique worth comparing against the other non-conventional techniques mentioned in Section 1.1. Extra heat-stable salts lower the pH of this sulfur-rich stream sent to a thermal reclamation unit, as shown in Figure 1. Here the addition of base (NaOH) raises pH and releases the protonated amine, which upon heating, volatilises the target amines leaving a caustic salt residue.

Here a similar lab method was applied to the CS-Cap SO_2 rich amine obtained from CSIRO's pilot campaign and the conditions at which the amine volatilised were investigated. Here after caustic treatment, the recovered amine (Monoethanolamine - MEA) was expected to vaporise near its boiling point (170°C). However, due to the high amount of heat-stable salts in the solution, there was a limited recovery of MEA under atmospheric conditions (atm pressure and 170°C). The high ionic strength (resulting from the large heat-stable salts burden and caustic added for neutralisation Wang et al., 2014) increases the boiling point of the solution sent for reclaiming. Hence, vacuum conditions of 50 mmHg and 110–130 °C were investigated to obtain higher amine recoveries. Further, the amount of caustic added was based on either the moles of heat-stable sulfate salt present or the pH of the final solution. Laboratory experiments were conducted on both a pilot-plant sample, which was initially ~3M MEA (~14 wt% amine after flue gas absorption), and synthetic sample, which was initially 5M MEA and reached ~ 25 wt% after being loaded with CO_2 and SO_4. Due to their high sulfate level, the amine solution sent for thermal reclamation has minimal CO_2. Thermal decomposition of MEA nominally occurs at temperatures >200° C, particularly in the presence of CO_2 , reclaimer temperatures of 150°C are known to result in thermal degradation. (Vega, et al. 2014; Gouedard, Picq, Launay, & Carrette, 2012; Dai et al., 2012). Oxazolidione is a reversible intramolecular condensation product of the MEA Carbamate and the first step in thermally induced carbamate polymerisation (Vega, et al. 2014). At the lower 110°C-130°C temperatures used for our high level costing, thermal degradation is considered less significant. No investigation was carried on the additional degradation products formed as a result of thermal reclamation process.

The details of the laboratory thermal reclaiming experiments are available in our previous publication (Garg et al., 2018b). For costing purposes, we built and validated an Aspen Plus simulation model using the laboratory experimental conditions. The laboratory-scale model was then was scaled up for a large-scale facility and equipment as shown in the thermal reclamation section in Figure 1. This included the addition of a packed contactor with a reboiler, mixing tank and pump. The amine vapour exiting the thermal reclaimer was added directly to the stripper section of the CO2-rich stream regeneration. Chemical compositions of the experimental samples were used to define the various input streams in the simulation model. After selecting the specific equilibrium model, property packages in Aspen Plus were used to simulate operating conditions. The results of the simulation were compared with experimental results to verify whether the selected equilibrium model and property package adequately replicated the actual experimental results. The stream flowrates were then scaled up in order to estimate the cost of a commercial facility. The model's predictions were evaluated against experimental results achieved with pilot and synthetic samples at various operating conditions. The details of the scale up are mentioned in our previous publication (Garg et al., 2019)

2.1.2. Option 2: Reactive crystallisation for regeneration of sulfur-rich amine stream

Crystallisation from a solution typically occurs as a result of cooling or evaporation, but can also occur via reaction of two solutes in a saturated solution. Though crystallisation is not commonly applied in gas processing or carbon capture, it is applicable to CS-Cap, due to the high concentration and ionic nature of the heat-stable salts in the sulfur-rich stream.

Figure 1 includes a schematic of reactive crystallisation (regeneration option 2) and its integration into the CS-Cap process. The addition of sufficient KOH to the sulfate-rich amine absorbent releases protonated amine as free amine, and the SO_4^{2-} reacts with K^+ to form K_2SO_4 crystals - a potential fertiliser. The effectiveness of crystallisation was previously evaluated at laboratory scale(Garg, et al., 2018b). The technique involved conditions close to those anticipated in the absorber to avoid any excessive cooling or heating load, with the aim being to draw out more sulfate under optimised experimental conditions.

The laboratory data was used to validate an Aspen Plus model, which was scaled up to allow costing of the technique at full scale. While laboratory-scale tests require only a simple reaction beaker, vacuum filtration and oven drying, at full scale the technique requires a crystalliser, hydro cyclones, centrifuge and a dryer. The methodology of the scale-up is as defined in section 2.1.1. The details of the scale up are mentioned in our previous publication (Garg et al., 2019)

2.2. Assumptions and economic inputs to regeneration cost estimation

2.2.1. Flowrates of sulfur-rich amine stream

Using ProTreat^{TM}, CSIRO has completed an in-house simulation of the SO_2 and CO_2 absorbers from the CS-Cap process applied to a 900-MWe (gross output) coal-fired power station (Cousins et al., 2019). The net power output after addition of 90% CO_2 capture using 30 wt% MEA as an absorbent was 634 MWe. The results of that simulation are used to provide the flowrate and composition of the sulfur-rich stream to be sent for reclamation. The simulations were completed for both brown (sulfur 200 ppm) and black (sulfur 700 ppm) coal-fired power stations.

2.2.2. Composition of sulfur-rich amine stream for regeneration

The full-scale system has three trains, the flowrate from one train was used for the simulation in Aspen Plus and formed the basis of the cost estimation. From the total flow rate of CO_2-rich amine, 0.1% (Beyad et al., 2014) of the stream was used as a bleed stream to capture SO_2 in the lower portion of the absorber. This stream was sized to remove the formate at the rate at which it formed. Hence the fraction of rich absorbent separated to the SO_2 loop varied between 0.1-0.15% .Of the SO_2-rich stream coming from the bottom section of the absorber column, 98% was recycled and the remaining 2% sent for the reclamation (Cousins et al., 2019).

A 30 wt% MEA solution was used to absorb the SO_2 and CO_2 in the simulation. Formate was included as a representative MEA oxidative degradation product given it is a major contributor to the product mix. MEA oxidation is more rapid than carbamate polymerisation (sometimes referred to as thermal degradation) (Vevelstad et al., 2013; Reynolds et al., 2015) and Oxazolidione is not a stable thermal degradation product (Gouedard et.al, 2012). The model was kept relatively simple by using formate as the only degradation product and it is an established marker for modelling oxidative degradation (Dhingra et.al, 2017). The total amine present after absorption was divided into protonated and free amine to match the final pH of the absorbent as obtained in the laboratory experiments. Table 1 shows the flow rate and composition of the stream used in Aspen Plus for both regeneration techniques.

Table 1: Composition and flow rates used for Aspen Plus simulation regeneration model for one post-combustion CO_2 capture train (900 MWe gross)

Component	**Thermal reclamation**		**Reactive crystallisation**	
	700 ppm sulfur mass fraction (kg/kg)	200 ppm sulfur mass fraction (kg/kg)	700 ppm sulfur mass fraction (kg/kg)	200 ppm sulfur mass fraction (kg/kg)
Water	0.62	0.5	0.57	0.48
Monoethanolamine (MEA)	0.07	0.2	0.06	0.18
CO_2	0.03	0.08	0.02	0.08
$MEAH^+$	0.1	0.1	0.1	0.11
SO_4^{2-}	0.17	0.095	0.16	0.09
K^+	0	0	0.07	0.04
$HCOO^-$	0.01	0.02	0.01	0.02
Sulfur-rich amine flowrate (tonne/hr)	29.1	14.4	30.5	14.6
Temperature (°C)	40.4	37.3	40.4	37.3

2.2.3. Capital cost

The capital cost was divided into direct plant costs, indirect costs and working capital. Only the equipment purchase cost (EPC) was taken from Aspen Plus. The remaining expenses were based on a percentage of the equipment price (Ulrich & Vasudevan, 2004), as outlined in Supplementary Information (SI) tables A, C, E and G for thermal reclamation and crystallisation, respectively.

2.2.4. Operating costs

The operating costs included the raw material and utilities costs as variable components. The labour charge, maintenance and repairs, taxes and insurance were included under fixed operating charges. Regeneration unit labour for O&M was factored in at AUD$73 per hour for 1800 hr p.a. Fixed capital depreciation and interest on capital was also included under operating costs. No administration, waste disposal cost, co-product or by-product credit were considered.

2.2.5. Utilities and chemical requirements

For thermal reclamation, low-pressure steam was used in the reboiler section. It is anticipated that this steam would be available at around 4.8 bar (APGT, 2015). An electrical load was considered for pumping. NaOH solution is required for the caustic treatment; for the Australian case, it was assumed that any pelletised NaOH used to prepare this solution would be imported. A MEA price of US$1.8/kg and a NaOH pellet price of US$385/tonne were used (CSIRO in-house chemical database). The price of steam, cooling water and electricity were taken from (Hosseini, Haque, Selomulya, & Zhang, 2016).

Cooling water is required to operate the crystalliser at a constant temperature. Electrical loads are also needed to run the hydro cyclones, centrifuges, dryers and pumps. Compressed air heating used direct contact electrical heaters. The solid KOH was assumed to be imported in a similar manner as the NaOH at a price of US$1200/tonne.

All capital and O&M costs in this paper are expressed in June 2015 Australian dollars similar to those considered in the reference reports (exchange rate of 1 AUD = 0.71 USD).

3. Cost comparison: CS-Cap process with conventional PCC technologies retrofitted to an Australian coal-fired power plant

The overall cost of the CS-Cap process is estimated, then compared to a conventional PCC + FGD configuration. As Australian coal-fired power plants do not have FGD units installed, our estimated CO2 capture cost includes the cost of FGD installation, whether for a new build or retrofitted power plant. The CS-Cap process cost is benchmarked against those reported previously for Australian coal-fired power plants using PCC + FGD. Reports prepared by Gamma Energy Technology and CO2CRC (CO2CRC, 2017) and Electric Power Research Institute (APGT, 2015) were used for cost comparison.

3.1. Assumptions, economic inputs and basis of reference reports for cost comparison

The retrofitted PCC facility evaluated in the (CO2CRC, 2017) study included two absorber trains, one regenerator per absorber train, one compression train per regenerator and eight reboilers per regenerator with CO_2 capture rate set at 90% with 24/7 operation. The PCC plant retrofit obtains steam for solvent regeneration via the crossover between intermediate-pressure and low-pressure systems in the existing plant. A backpressure steam turbine is introduced to step down the steam to the correct conditions for solvent stripping in the reboiler. The heat from the hot condensate returning from the reboiler supplements feedwater heating via heat exchangers. This base case PCC coal retrofit was a fully integrated MEA solvent facility. For the retrofit case, the existing pulverised-coal base plant without capture is a fully paid-off asset, in good condition with a suitably long life

(CO2CRC, 2017). The retrofit case assumes wet cooling and no upgrades are considered to the base plant.

Another report (APGT, 2015) details the cost and performance of black and brown-coal supercritical and ultra-supercritical coal-fired power plants with and without CCS. It includes the costs of FGD installation and of retrofitting PCC without FGD in an existing Australian black-coal-fired power plant. This information is used here to compare the CS-Cap case with other reference cases using a standard CO_2 capture process.

3.2. Scenarios comparing the CS-Cap process with conventional PCC + FGD installations

The following scenarios are detailed in SI:

- Scenario 1: Base case retrofitted with PCC, FGD + selective catalytic reduction (SCR)
- Scenario 2: Base case (with FGD & SCR and improved solvent)
- Scenario 3: Base case (with minimal FGD & SCR)
- Scenario 4: CS-Cap case with thermal reclamation
- Scenario 5: CS-Cap case with reactive crystallisation

3.3. Calculation basis for Total Plant Costs (TPC) of scenarios 1–5

The TPC in A$/kW for scenarios 1–3 with wet cooling (as applicable to black and brown coal) are taken from the (CO2CRC, 2017) report. Any PCC deployment cost in Australia has two components: CO_2 capture (PCC cost) and FGD cost for SO_2 capture. The cost breakup of TPC for PCC Retrofit + FGD/SCR with dry cooling is provided in (APGT, 2015). The TPC of PCC equipment retrofit + FGD/SCR for a wet cooling case is available (CO2CRC, 2017), however, no breakdown is available for the individual PCC retrofit and FGD/SCR costs.

To evaluate the PCC retrofit TPC in scenarios 4–5, the ratio of FGD/SCR cost requirements to the total TPC using dry cooling for a black-coal power plant is used to calculate the individual PCC equipment retrofit cost. An air-cooled condensing plant design can reduce plant gross output by as much as 1% compared with wet cooling, however, a similar cost ratio of 44–45% for FGD/SCR (dry cooling) is used here. The cost obtained for FGD/SCR was then subtracted from the combined TPC of the Scenario 1 PCC retrofit + FGD/SCR. The cost of the two CS-Cap regeneration techniques is then added to this modified TPC. This replaces the cost of FGD/SCR with the cost of the CS-Cap process.

3.4. Calculation basis for fixed and variable O&M costs for scenarios 4–5

The total fixed and variable costs in A$/kW-yr and A$/MWh for scenarios 1–3 have been taken from the report (CO2CRC, 2017). The individual fixed cost breakdown for PCC retrofit and FGD/SCR, used a similar factor of 44–45% as identified in (APGT, 2015). This is based on the cost estimation method, which assumes that the fixed part of the operating cost is taken as a percentage of the capital cost. By excluding FGD, the resulting 45% reduction in capital cost afforded the same reduction in fixed operating cost. The fixed O&M cost of regeneration techniques was then added to the fixed O&M cost of the PCC retrofit obtained from the report.

In the (APGT, 2015), FGD/SCR was found to provide 20% of the variable O&M costs. This factor was applied to the variable O&M costs obtained from the (CO2CRC, 2017). This allowed the FGD/SCR portion of the variable O&M costs to be replaced by those calculated for the CS-Cap process.

The TPC, operating and maintenance costs as outlined in SI Tables A, B, E and F have been calculated for a 900-MW gross power plant. However, the costs provided in (APGT, 2015) for the black-coal cases are calculated based on a 450-MW gross power plant with a net output of 331 MW with PCC. Hence, all costs mentioned in Tables A, B, E and F are converted to a similar basis for a 450-MW power plant, as shown in Table I (SI), to provide a similar basis for all scenarios to be compared. A power law equation commonly used to scale plant and equipment costs with changes in throughput or capacity has been used to adjust the costs (Haque & Somerville, 2013):

$$\text{New plant cost} \;=\; \text{Base Cost X (New capacity/base capacity)}^{\,0.65} \quad (1)$$

3.5. Calculation basis for LCOE & cost of CO_2 avoided

The LCOE is the average cost of producing electricity from any technology over its entire life. It is calculated by converting the capital and O&M costs of a plant into a cost per MWh.

The cost of electricity consists of four components: capital costs, O&M costs, fuel costs & CO_2 transportation and sequestration. When these costs are calculated independently, individual cost basis comparison can be misleading. To compare different size units, they must all have the same cost unit basis when combined to calculate the cost of electricity (typically $/MWh).

The criteria used to convert all costs including the capital, O&M costs into $/MWh is described in Section 17.2.5 of (APGT, 2015). In addition, the Average LCOE and cost of CO_2 avoided has been calculated as follows:

$$\begin{aligned} Average\ LCOE &= Finance\ Charges + Fixed\ O\&M + Variable\ Cost + Fuel\ Cost \\ &\quad + Cost\ of\ Transportation\ \&\ Storage \end{aligned}$$

(~~1~~2)

$$\frac{Cost}{\mathrm{MWh}} = \frac{Cost\ \text{in AUD per kW} \times\ Sent\ Output(kW)}{Sent\ Output(kW) \times Operation\ Hours\ \times\ Capacity factor} \times 1000$$

(~~2~~3)

All the various costs fixed or variable need to converted in to Cost/MWh to evaluate average LCOE.

Operation hours considered = 8760 per year,

Capacity factor=0.85

$$Avoided\ cost = \frac{\{LCOE_{with\ removal} - LCOE_{reference}\}\$/MWh}{\{CO2\ emissions_{reference} - CO2\ emissions_{with\ removal}\}tonnes/MWh} \quad (4)$$

4. Results and discussion

4.1. Cost estimation results of regeneration options 1 and 2

As the incoming sulfur content in flue gases affects the capital and operating cost of the reclaimers, regeneration options 1 and 2 have been evaluated for both high and low-sulfur-content gases. Table 2 summarises the total costs for each scenario.

Supplementary Tables A and B detail the capital expenditure and operating costs of thermal reclamation when applied to the CS-Cap process for 700 ppm while Tables C and D cover 200ppm sulfur content in the incoming flue gas.

Supplementary Tables E and F detail the capital expenditure and operating cost of reactive crystallisation when applied to the CS-Cap process for 700 ppm while Tables G and H cover 200ppm sulfur content in the incoming flue gas.

Table 2: Total cost of reclamation techniques at 200ppm and 700ppm incoming sulfur content in flue gas

Description	Flue gas Sulfur (ppm)	Total Cost in AUD ex taxes were applicable	
		Capital($M)	Operating($M/yr)
Option 1 -Thermal reclaimer for one train of 900 MW plant	700	4.25	71.73
	200	2.26	37.59
Option 2 - Reactive crystalliser for one train of 900 MW plant	700	7.26	59.89
	200	7.05	26.19

Comparing costs outlined in Table 2, it can be seen that crystallisation is a cheaper regeneration technique than thermal reclamation for the CS-Cap process for any level of sulfur in the flue gas. The crystallisation capital requirements are nearly 1.7 times those of thermal reclamation due to the multiple process operations involved (supplementary Tables E-H). However, this cost is offset by the comparatively lower MEA losses in crystallisation which are approximately 0.5-0.6 times to that of thermal reclamation.

Unlike reactive crystallisation, which will not remove non-ionic degradation products, thermal reclamation is capable of removing most impurities. Due to the slow build-up of degradation products (apart from sulfate), periodic thermal reclaiming is likely still required in addition to crystallisation to remove non-ionic impurities. This research focused only on the removal of sulfate; hence, the comparison is based only on the efficiencies of sulfate removal from the spent absorbent.

Cost comparisons are provided next for a black coal fired plant (Table 3), then brown coal (Table 4) and both are summarised on a LCOE basis (Figure 2).

4.2. Cost comparison results of scenarios 1–5 (black-coal case)

To retrofit PCC to existing Australian coal-fired power plant, the total plant cost for CO_2 capture will include the TPC of PCC + FGD/SCR. This is applicable for scenarios 1–3 of Section 3. However, for CS-Cap scenarios 4–5, the TPC will include PCC retrofit + TPC of reclaimer (either thermal reclaimer or reactive crystalliser).

4.2.1. PCC retrofit cost & FGD/SCR cost capital cost

The cost of the regeneration techniques outlined in options 1 and 2 and calculated in Section 4.1 is for one train of the PCC system installed at a 900-MW gross power station. The net output of the plant is 662 MWe for a black-coal case.

Further, for CS-Cap case scenarios 4 and 5, the basic PCC equipment retrofit cost is slightly higher than scenarios 1–3. This is because the CS-Cap absorber column has to be larger than a standard column as a result of the additional SO_2 absorption section, which requires an additional packing height of 4 m. The CO_2 capture section is not altered, as the increased flue gas flow rate through the blower (depending on its location) and cooling duty applied to the direct contact cooler are only slight. The water balance in the SO_2 capture loop can be maintained by operating the column close to 40 °C.

Table 3: Total plant costs for retrofitting post-combustion capture (PCC) in a 450-MW (331 MW net) black-coal power plant

Cost	**Scenario 1**	**Scenario 2**	**Scenario 4**	**Scenario 5**
	Base case (with FGD & SCR)	Base case (with FGD & SCR and improved solvent)	CS-Cap case with thermal reclaimer	CS-Cap Case with reactive crystalliser
PCC equipment retrofit cost ($/kWe)	2,283	2,118	2,285	2,285
FGD + SCR retrofit cost ($/kWe)	1,817	1,742	24	42
Total plant cost (A$/kW sent out)	4,100	3,860	2,309	2,327
Total plant cost (A$bn)	1.9	1.4	1.0	1.0

FGD = flue gas desulfurisation; SCR = selective catalytic reduction

Table 3 enables the TPC comparison of scenarios 1–5 for the black-coal case. Scenario 3, i.e. PCC with minimal FGD/SCR, was not considered in the reference report and is not included here. Table 3 reveals that the capital cost of retrofitting the CS-Cap process, using a thermal reclaimer or crystalliser in an existing plant, would be the cheapest option for an Australian black-coal-fired plant. Within the accuracy of these cost estimates, the total plant costs for scenario 4 and 5 were found to be the same at A $1bn. As mentioned previously, the crystalliser will not remove all potential impurities, and hence periodic thermal reclamation may be needed, which will slightly raise the cost of Scenario 5. On the other hand, Scenario 4 does not need any additional batch thermal reclaimer, as the continuous thermal reclaimer of the CS-Cap process will be able to remove degradation products other than sulfate. The TPC reduction from Scenario 1 to Scenario 4 is around 43%. Note that scenarios 4 and 5 do not include any SCR installations. This is because the NOx component of the flue gas from Australian coal power plants is mostly “inert” NO (~99% NO, balance NO2 and N2O) (Meuleman, et al. 2010). NO does not react with amines, and hence passes through without affecting the CO_2 capture system. The FGD/SCR retrofit cost reductions are considerable when comparing Scenario 1 to scenarios 4 and 5.

The capital cost comparison in Table 3 definitely favours the CS-Cap process for Australian black-coal-fired power plants. However, it is important to calculate the LCOE, which includes the O&M, fuel, CO_2 transportation and sequestration costs as listed in SI tables K and N. This will identify whether the electricity produced could be cheaper, against standard technologies, if the CS-Cap process is used.

4.2.2. LCOE & CO_2-avoided cost for scenarios 1–5

Table J (SI) reveals the capture costs for retrofitting PCC to a black-coal power plant for scenarios 1–5. The data for scenarios 1 and 2 is taken from the reference report (CO2CRC, 2017). The various cost data for scenarios 4 and 5 have been calculated based on the method described in sections 3.2–3.4 and the data calculated in SI Table I. These retrofit costs in SI Table J are used to calculate the LCOE for our black-coal case (SI Table K). The average LCOE data for black coal case are summarised in Figure 2. Figure 2 reveals that the cost of electricity is comparable between the base cases and retrofitting either CS-Cap process.

4.3. Cost comparison results for scenarios 1–5 of Section 3 (brown-coal case)

4.3.1. PCC retrofit cost and FGD/SCR cost

Scenarios 1–2 are based on a supercritical 2100-MW brown-coal power plant with a net output of 1283 MW when retrofitted with PCC. Hence, for LCOE cost comparison, the costs of amine regeneration were converted to the larger plant size.

Further, the capital and operating costs for the thermal reclaimer and reactive crystalliser are calculated in SI tables C, D, G and H (200ppm Sulfur).The cooled flue gas flow rates for the brown and black coal cases are anticipated to be comparable for a similar capacity power plant. However, the brown coal power plant efficiency would reduce. A net capacity of 550 MW is assumed here for a 900-MW gross output brown-coal power plant for scenarios 4 and 5.

Table L (SI) shows the TPC and O&M costs adjusted for reclamation techniques applied to a 2100 MW (1283 MW net) brown-coal power plant for scenarios 1–5. The capital cost comparison in Table 4 again favours using the CS-Cap process for Australian brown-coal-fired power plants. Within the accuracy of these cost estimates, the total plant costs for scenario 4 and 5 were again found to be the same at A $5.8bn. However, the LCOE was calculated further below to identify whether the electricity produced would be cheaper against standard techniques if the CS-Cap process is used in Australian brown-coal-fired power plants.

Table 4: Total plant costs for retrofitting post-combustion capture (PCC) in a 2100-MW (1283 MW net) brown-coal power plant

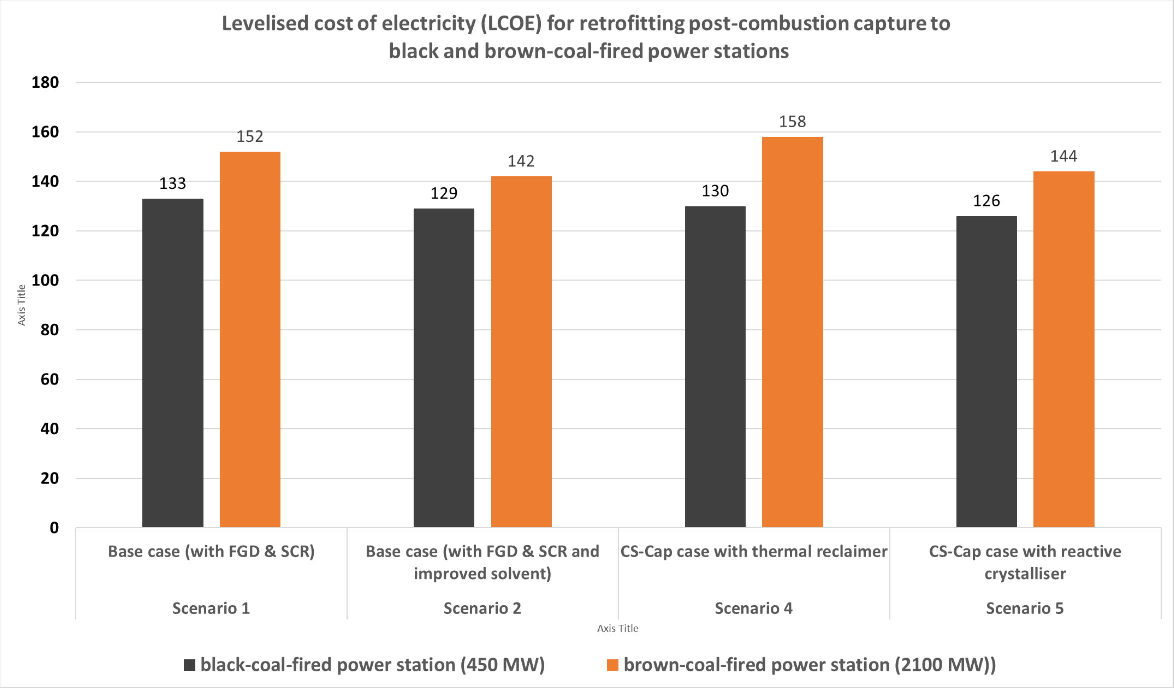
Cost	**Scenario 1**	**Scenario 2**	**Scenario 3**	**Scenario 4**	**Scenario 5**
	Base case (with FGD & SCR)	Base case (with FGD & SCR and improved solvent)	Base case (with minimal FGD & SCR), single boiler retrofit	CS-Cap case with thermal reclaimer	CS-Cap case with reactive crystalliser
PCC equipment retrofit cost ($/kWe)	2,728	2,579	2,728	2,730	2,730
FGD + SCR retrofit cost ($/kWe)	2,172	2,121	1172	9	29
Total plant cost (A$/kW sent out)	4,900	4,700	3,900	2,739	2,759
Total plant cost (A$bn)	10.3	9.9	1.5	5.8	5.8

FGD = flue gas desulfurisation; SCR = selective catalytic reduction

4.3.2. LCOE & CO_2-avoided costs for black and brown coals in scenarios 1–5

Table M (SI) shows the capture costs for retrofitting PCC to the brown-coal case for scenarios 1–5. The data for scenarios 1,2 and 3 is taken from the reference report (CO2CRC, 2017). The costs in (SI) Table M are used to calculate the LCOE for the brown-coal case (SI Table N) and are summarised here in Figure 2. For comparison, Figure 2 also includes the LCOE results for black coal scenarios discussed previously in section 4.2.2.

Figure2: - Levelised cost of electricity (LCOE) for retrofitting post-combustion capture to black and brown-coal-fired power stations



FGD = flue gas desulfurisation; SCR = selective catalytic reduction. For LCOE comparison, the calculated LCOE without CCS is $39/MWh with CO_2 emissions of 2.95 Mt/year for a Black coal 450-MW power plant. For a brown coal plant, the LCOE without CCS is $29/MWh with CO_2 emissions of 19.88 Mt/year for 2100-MW. Figure 2 reveals that the lowest cost of electricity is achieved by retrofitting the CS-Cap process using a reactive crystalliser to a 450-MW (331 MW net) black-coal power plant.

Further, in Figure 2, when comparing scenarios 1, 4 and 5, CS-Cap is also the cheapest option if used with a crystalliser in a brown-coal-fired power plant. The LCOE for scenario 2 and 3 is calculated only for single boiler retrofit (343 and 380 MW respectively) whereas scenario 1, 4 and 5 are calculated for full capacity (1283 MW) retrofit. The cost of retrofitting FGD+PCC reduces further if only a single boiler is retrofitted, as shown in scenarios 2 and 3. However, comparing CS-Cap cases for scenarios 4 and 5 against scenarios 2 and 3 is not valid. The CS-Cap cases are not evaluated here for single boiler retrofit, which will further reduce the cost of CS-Cap scenarios 4 and 5. The case of minimal FGD/SCR with a single boiler retrofit (Scenario 3) results in the lowest LCOE for the brown-coal case, as mentioned in the report (CO2CRC, 2017). This optimistic approach of applying minimal FGD/SCR as reported in(CO2CRC, 2017) is the cheapest option for Australian coal plants, but this theoretical outcome was not based on any bench or pilot-scale study. In contrast, the present CS-Cap costing study is based on the results of pilot and bench-scale testing and is considered more reliable. In addition, the potential cost benefit from selling by-products has not been included in this costing study. It might be possible to improve the economics of CS-Cap by selling K_2SO_4 generated as a by-product of the process.

Estimations of the overall CO_2 avoidance cost for both coal types are presented in Table 5

Table 5: Summary Table –CO_2 avoided cost for retrofitting post-combustion capture to black and brown-coal-fired power stations

Cost CO_2 avoided ($/tonne)	**Scenario 1** Base case (with FGD & SCR)	**Scenario 2** Base case (with FGD & SCR and improved solvent)	**Scenario 4** CS-Cap case with thermal reclaimer	**Scenario 5** CS-Cap case with reactive crystalliser
black-coal-fired power station (450 MW)	**125**	**117**	**121**	**116**
brown-coal-fired power station (2100 MW)	**120**	**104**	**126**	**112**

Shifting the focus to modelling the cost of flue gas CO_2 removal instead of electricity reveals (Table 5) that only the regeneration using recrystallization (Scenario 5) produced lower values than the base case. CS-Cap incorporating reactive crystallisation is worthy of further investigation given its calculations are based on using the higher energy demand MEA solvent.

5. Conclusions and future work

High level modelling estimated the costs for the conceptual CS-Cap PCC process, with a focus on two alternative solvent recovery technologies. Both of the technologies evaluated experimentally in this paper can regenerate the sulfate-rich aqueous amine absorbent produced by the CS-Cap process. Due to its lower amine makeup requirements, modelling CS-Cap with reactive crystallisation revealed consistently lower LCOE and CO2 cost avoided values compared to thermal reclamation.

We found minimal differences in LCOE for a black-coal power plant comparing base case PCC ($133/MWh) with FGD and SCR against CS-Cap with a reactive crystalliser ($126/MWh) for both new build and retrofitting to an existing (450 MW) power plant. We can conclude that the CS-Cap process is unlikely to offer significant savings to Australian black-coal-fired power plants. The key factor being their higher SO_2 emissions compared to brown coal plants.

For a brown-coal 2100 MW power plant, the calculated LCOE is slightly lower for CS-Cap process using a crystalliser ($144/MWh) compared with a standard PCC retrofit process with FGD ($152/MWh).

The greatest contributor to the LCOE of CS-Cap is operating cost, rather than capital cost, due to its continuous and relatively high chemical requirements such as amine solvent makeup and alkali (KOH or NaOH) addition. The operating cost of regeneration increases by nearly 50% if sulfur levels rise from 200 to 700 ppm in the flue gases, e.g. resulting in an additional MEA loss of approx. 14% and an increase of 17-18% in CO_2 avoidance cost. For the lower sulfur brown coal, amine losses of ~15% for thermal reclamation and ~8–9% for reactive crystallisation are a significant operational cost.

Future work should therefore focus on further improving data quality and reducing operating costs by conducting experiments at pilot scale on brown coal plant.

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Highlights.

Title: Techno-economic evaluation of amine-reclamation technologies and combined CO2/SO2 capture for Australian coal-fired plants: manuscript JGGC_2019_643

- CS-Cap technology has previously shown potential as a means of reducing the cost of PCC and suited to retrofitting legacy power plants
- This paper is the first to develop a cost model of the CS-Cap technology based on actual laboratory data. This data is combined with CSIRO’s legacy pilot-scale results and peer reviewed costings for related conventional PCC technologies and Australian power plants.
- The key variable impacting costs was found to be solvent regeneration
- The cost reductions were not found to be as significant as expected
- Flue gas sulfur was found to be the main influence on regeneration costs

Techno-economic evaluation of amine-reclamation technologies and combined CO_2/SO_2 capture for Australian coal-fired plants

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Abstract

CSIRO's patented CS-Cap process aims at reducing the costs of amine-based post-combustion capture by combining SO_2 and CO_2 capture using one absorbent in a single absorber column. By avoiding the need for a separate flue gas desulfurization unit, the process offers potential savings for power plants requiring CO_2 capture. High-level cost estimates based on lab and pilot data are presented for two amine reclamation techniques i.e. thermal reclamation and reactive crystallisation. Only regeneration via reactive crystallisation reduces CS-Cap costs below base case FGD/SCR-PCC. Cost estimations suggest a potential reduction of 38–44% in the total plant cost when using the CS-Cap process compared to base case. However, the amine reclaimer operating cost governs the overall cost of the CS-Cap process and is highly sensitive to sulfur content. A 50% reduction is observed when SO_2 levels reduce from 700 to 200 ppm. Comparing levelised cost of electricity and CO_2 avoided costs for CS-Cap against our base case, low sulfur brown coal has a slight (5-7%) cost advantage; however, confirmation requires pilot data on amine recovery.

Keywords: CS-Cap; thermal reclamation; reactive crystallisation; total plant cost; Aspen Plus Economic Analyser; cost estimation

1. Introduction

Approximately 70% of Australia's electricity generation comes from local black and brown coal (Engineers Australia, 2017). However, coal-based electricity generation is facing significant emission constraints due to the need for climate change mitigation.

Future energy generation and security will not be solely dependent on any one power technology. Instead, a strategic approach that combines existing generation with emerging low-emission technologies will lead to a smooth and sustainable transition. As per the Australian Government's Energy White Paper (Engineers Australia, 2017), coal could continue supplying energy, provided emissions are reduced at low cost. Integrating carbon capture and storage (CCS) into power stations will significantly reduce emissions, but will result in high-cost electricity. In the net zero CO_2 by 2050 scenario, CCS needs fast deployment and that requires lower-cost capture technologies. This paper investigates such a technology i.e. a single column SO_2 and CO_2 capture process, and estimates the cost of key process and the amine regeneration scenarios.

1.1. CSIRO's CS-Cap process

Amine-based Post Combustion Capture (PCC) is currently the leading technology for capturing CO_2 from power plants (IEAGHG, 2019). Pre-treatment of flue gases, including removal of SO_2, is essential for efficient PCC. SO_2 preferentially reacts with amines to form degradation products, thus increasing amine makeup requirements (CO2CRC, 2017). The prior removal of SO_2, requires flue gas desulfurisation (FGD). In Northern Hemisphere plants, FGD is typically conducted via a wet limestone process (Jamil et.al, 2013). Amine based PCC requires FGD units to reduce SO_2 levels below 10 ppm to minimise degradation. However, existing FGD units may not reduce SO_2 levels to such low levels, and these plants require additional scrubbers or deep FGD units (Puxty et al., 2014). Australian plants do not use FGD units (Puyvelde, 2009), therefore, any PCC retrofit requires additional investment (Rubin, Davison, & Herzog, 2015).

As an alternative to high-cost FGD or upgraded FGD, CSIRO has developed a combined SO_2 + CO_2 capture process utilising a single aqueous amine solution in a single absorber column (Beyad et al., 2014). Known as CS-Cap, the process offers potential capex savings by avoiding the need for a separate FGD unit (Puxty et al., 2014). A proof of concept operation was conducted at CSIRO's pilot facility, at AGL's Loy Yang brown-coal-fired power station (Pearson et al., 2017). A flow diagram of the CS-Cap process, which has been patented by CSIRO, is provided in SI Figure A, the absorber captures CO_2 and SO_2 in upper and lower sections of the absorber column, respectively. The selectivity of amines to absorb stronger acids allows SO_2 to be absorbed in the bleed stream despite the presence of CO_2. Only a small percentage of the total volumetric flowrate (~0.01–3%; Beyad et al., 2014) of the CO_2-rich stream (red stream) is required to capture the incoming SO_2 in the flue gas. The CO_2-rich amine from the top section of the absorber column is regenerated by a standard steam stripping process. However, it is not possible to thermally regenerate the SO_2-rich amine formed in the lower section of the absorber as it forms a heat-stable sulfate (Pearson et al., 2017). Strong ionic bonding between protonated amines and sulfate ensures these salts do not revert to amine and SO_2 under the optimal heat treatment conditions for CO_2 release. Hence, a part of the recycle stream exiting from the bottom section of the absorber column is sent to a separate regeneration section, which is the sulfur-rich amine regeneration unit (Figure A).

The composition of the sulfur-rich CS-Cap stream sent for amine regeneration is different from a typical reclaimer stream, given its much higher (11–12 wt% versus 2 wt%) heat-stable salt burden (Garg et al., 2018). Here, the regeneration of amine can occur through thermal reclamation (ElMoudir, Supap, & Saiwan, 2012) but also through non-conventional regeneration methods.

Alternative techniques such as electrodialysis, ion exchange, nanofiltration and crystallisation are capable of regenerating amines (Garg et al., 2018) from their sulfate salts.

1.2. Investigating the cost effectiveness of the CS-Cap process

Adding CCS to a coal-fired power plant doubles the total plant cost (TPC) regardless of coal type (APGT, 2015). The variable operation and maintenance (O&M) cost is expected to rise even higher (around 3–4 times) if CCS is integrated into a new power plant. This increases the levelised cost of electricity (LCOE) from a pulverised supercritical or ultra-supercritical power plant by approximately 100% (APGT, 2015). This cost includes FGD unit installations, which as mentioned above, are a pre-requisite for any amine-based PCC technology deployed in Australia.

The huge increase in the capital and O&M costs of a new power plant built with CCS increases electricity prices while decreasing plant efficiency. This poses a serious barrier to CCS commercialisation in Australia. As described above, CSIRO’s CS-Cap technology aims at reducing the capital cost of amine-based PCC by eliminating the capital cost of a separate FGD unit. However, the cost effectiveness of regenerating the CO_2 and SO_2-rich amine still needs to be investigated and compared with a standard FGD+PCC system.

As shown in Figure A (SI), amine regeneration is chiefly required for two streams in the CS-Cap process: the CO_2-rich amine stream (shown in red) and the SO_2 rich amine stream (shown in green). The CO_2-rich stream is regenerated in a standard stripper using steam. The cost of regenerating the sulfur-rich amine stream has scope for further reduction to that afforded by thermal reclamation. This is due to its unique nature (highly concentrated with sufur in form of sulfate), making it applicable to alternative lower energy reclamation techniques (Garg et al., 2018b).

The practicality and cost effectiveness of thermal reclamation and crystallisation based regeneration technologies are experimentally evaluated here using a sulfur-rich CS-Cap absorbent. This paper includes a cost estimation of these regeneration techniques, using an Aspen Plus simulation, validated against the results of our lab experiments. The cost of these regeneration techniques is integrated individually with the unit operations shown in Figure A (SI) to evaluate the overall process. The cost of CS-Cap is compared with the cost of a power plant retrofitted with:

- FGD + PCC
- FGD + PCC + improved solvent
- Minimal FGD + PCC.

2. Cost evaluation method: CS-Cap regeneration techniques

2.1. Process description and model development

Figure 1 represents the CS-Cap process with the two selected options for sulfur-rich amine regeneration. Options 1 and 2 show the thermal reclamation and reactive crystallisation units, respectively. Each option was separately integrated with the combined absorption section (SO_2 and CO_2) and CO_2-regeneration section to determine the overall cost of the CS-Cap process.

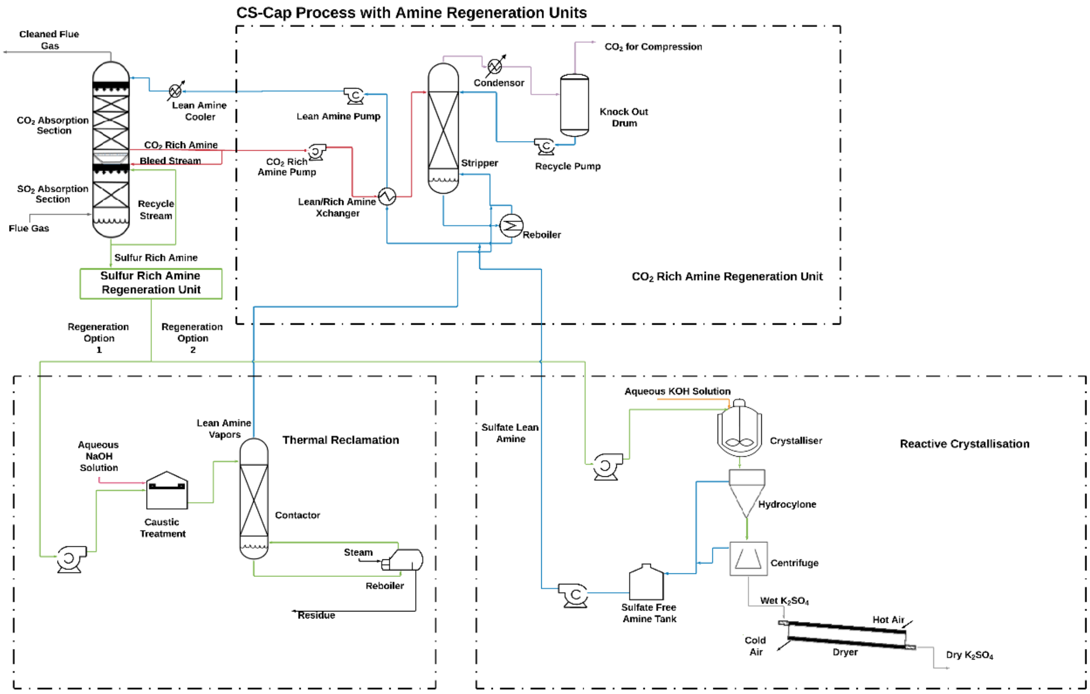


Figure 1: CS-Cap process showing reclamation of sulfur-rich amine through thermal reclamation or crystallisation

2.1.1. Option 1: Thermal reclamation

Compared to conventional HSS loadings, the more concentrated CS-Cap stream could require less or a similar amount of energy to recover the amine (Sexton et al., 2014, Wang, Hovland, & Jens, 2014). This makes the thermal reclaiming technique worth comparing against the other non-conventional techniques mentioned in Section 1.1. Extra heat-stable salts lower the pH of this sulfur-rich stream sent to a thermal reclamation unit, as shown in Figure 1. Here the addition of base (NaOH) raises pH and releases the protonated amine, which upon heating, volatilises the target amines leaving a caustic salt residue.

Here a similar lab method was applied to the CS-Cap SO_2 rich amine obtained from CSIRO's pilot campaign and the conditions at which the amine volatilised were investigated. Here after caustic treatment, the recovered amine (Monoethanolamine - MEA) was expected to vaporise near its boiling point (170°C). However, due to the high amount of heat-stable salts in the solution, there was a limited recovery of MEA under atmospheric conditions (atm pressure and 170°C). The high ionic strength (resulting from the large heat-stable salts burden and caustic added for neutralisation Wang et al., 2014) increases the boiling point of the solution sent for reclaiming. Hence, vacuum conditions of 50 mmHg and 110–130 °C were investigated to obtain higher amine recoveries. Further, the amount of caustic added was based on either the moles of heat-stable sulfate salt present or the pH of the final solution. Laboratory experiments were conducted on both a pilot-plant sample, which was initially ~3M MEA (~14 wt% amine after flue gas absorption), and synthetic sample, which was initially 5M MEA and reached ~ 25 wt% after being loaded with CO_2 and SO_4. Due to their high sulfate level, the amine solution sent for thermal reclamation has minimal CO_2. Thermal decomposition of MEA nominally occurs at temperatures >200^o C, particularly in the presence of CO_2 , reclaimer temperatures of 150°C are known to result in thermal degradation. (Vega, et al. 2014; Gouedard, Picq, Launay, & Carrette, 2012; Dai et al., 2012). Oxazolidione is a reversible intramolecular condensation product of the MEA Carbamate and the first step in thermally induced carbamate polymerisation (Vega, et al. 2014). At the lower 110°C-130°C temperatures used for our high level costing, thermal degradation is considered less significant. No investigation was carried on the additional degradation products formed as a result of thermal reclamation process.

The details of the laboratory thermal reclaiming experiments are available in our previous publication (Garg et al., 2018b). For costing purposes, we built and validated an Aspen Plus simulation model using the laboratory experimental conditions. The laboratory-scale model was then was scaled up for a large-scale facility and equipment as shown in the thermal reclamation section in Figure 1. This included the addition of a packed contactor with a reboiler, mixing tank and pump. The amine vapour exiting the thermal reclaimer was added directly to the stripper section of the CO2-rich stream regeneration. Chemical compositions of the experimental samples were used to define the various input streams in the simulation model. After selecting the specific equilibrium model, property packages in Aspen Plus were used to simulate operating conditions. The results of the simulation were compared with experimental results to verify whether the selected equilibrium model and property package adequately replicated the actual experimental results. The stream flowrates were then scaled up in order to estimate the cost of a commercial facility. The model's predictions were evaluated against experimental results achieved with pilot and synthetic samples at various operating conditions. The details of the scale up are mentioned in our previous publication (Garg et al., 2019)

2.1.2. Option 2: Reactive crystallisation for regeneration of sulfur-rich amine stream

Crystallisation from a solution typically occurs as a result of cooling or evaporation, but can also occur via reaction of two solutes in a saturated solution. Though crystallisation is not commonly applied in gas processing or carbon capture, it is applicable to CS-Cap, due to the high concentration and ionic nature of the heat-stable salts in the sulfur-rich stream.

Figure 1 includes a schematic of reactive crystallisation (regeneration option 2) and its integration into the CS-Cap process. The addition of sufficient KOH to the sulfate-rich amine absorbent releases protonated amine as free amine, and the SO_4^{2-} reacts with K^+ to form K_2SO_4 crystals - a potential fertiliser. The effectiveness of crystallisation was previously evaluated at laboratory scale(Garg, et al., 2018b). The technique involved conditions close to those anticipated in the absorber to avoid any excessive cooling or heating load, with the aim being to draw out more sulfate under optimised experimental conditions.

The laboratory data was used to validate an Aspen Plus model, which was scaled up to allow costing of the technique at full scale. While laboratory-scale tests require only a simple reaction beaker, vacuum filtration and oven drying, at full scale the technique requires a crystalliser, hydro cyclones, centrifuge and a dryer. The methodology of the scale-up is as defined in section 2.1.1. The details of the scale up are mentioned in our previous publication (Garg et al., 2019)

2.2. Assumptions and economic inputs to regeneration cost estimation

2.2.1. Flowrates of sulfur-rich amine stream

Using ProTreat^{TM}, CSIRO has completed an in-house simulation of the SO_2 and CO_2 absorbers from the CS-Cap process applied to a 900-MWe (gross output) coal-fired power station (Cousins et al., 2019). The net power output after addition of 90% CO_2 capture using 30 wt% MEA as an absorbent was 634 MWe. The results of that simulation are used to provide the flowrate and composition of the sulfur-rich stream to be sent for reclamation. The simulations were completed for both brown (sulfur 200 ppm) and black (sulfur 700 ppm) coal-fired power stations.

2.2.2. Composition of sulfur-rich amine stream for regeneration

The full-scale system has three trains, the flowrate from one train was used for the simulation in Aspen Plus and formed the basis of the cost estimation. From the total flow rate of CO_2-rich amine, 0.1% (Beyad et al., 2014) of the stream was used as a bleed stream to capture SO_2 in the lower portion of the absorber. This stream was sized to remove the formate at the rate at which it formed. Hence the fraction of rich absorbent separated to the SO_2 loop varied between 0.1-0.15% .Of the SO_2-rich stream coming from the bottom section of the absorber column, 98% was recycled and the remaining 2% sent for the reclamation (Cousins et al., 2019).

A 30 wt% MEA solution was used to absorb the SO_2 and CO_2 in the simulation. Formate was included as a representative MEA oxidative degradation product given it is a major contributor to the product mix. MEA oxidation is more rapid than carbamate polymerisation (sometimes referred to as thermal degradation) (Vevelstad et al., 2013; Reynolds et al., 2015) and Oxazolidione is not a stable thermal degradation product (Gouedard et.al, 2012). The model was kept relatively simple by using formate as the only degradation product and it is an established marker for modelling oxidative degradation (Dhingra et.al, 2017). The total amine present after absorption was divided into protonated and free amine to match the final pH of the absorbent as obtained in the laboratory experiments. Table 1 shows the flow rate and composition of the stream used in Aspen Plus for both regeneration techniques.

Table 1: Composition and flow rates used for Aspen Plus simulation regeneration model for one post-combustion CO_2 capture train (900 MWe gross)

Component	**Thermal reclamation**		**Reactive crystallisation**	
	700 ppm sulfur mass fraction (kg/kg)	200 ppm sulfur mass fraction (kg/kg)	700 ppm sulfur mass fraction (kg/kg)	200 ppm sulfur mass fraction (kg/kg)
Water	0.62	0.5	0.57	0.48
Monoethanolamine (MEA)	0.07	0.2	0.06	0.18
CO_2	0.03	0.08	0.02	0.08
$MEAH^+$	0.1	0.1	0.1	0.11
SO_4^{2-}	0.17	0.095	0.16	0.09
K^+	0	0	0.07	0.04
$HCOO^-$	0.01	0.02	0.01	0.02
Sulfur-rich amine flowrate (tonne/hr)	29.1	14.4	30.5	14.6
Temperature (°C)	40.4	37.3	40.4	37.3

2.2.3. Capital cost

The capital cost was divided into direct plant costs, indirect costs and working capital. Only the equipment purchase cost (EPC) was taken from Aspen Plus. The remaining expenses were based on a percentage of the equipment price (Ulrich & Vasudevan, 2004), as outlined in Supplementary Information (SI) tables A, C, E and G for thermal reclamation and crystallisation, respectively.

2.2.4. Operating costs

The operating costs included the raw material and utilities costs as variable components. The labour charge, maintenance and repairs, taxes and insurance were included under fixed operating charges. Regeneration unit labour for O&M was factored in at AUD$73 per hour for 1800 hr p.a. Fixed capital depreciation and interest on capital was also included under operating costs. No administration, waste disposal cost, co-product or by-product credit were considered.

2.2.5. Utilities and chemical requirements

For thermal reclamation, low-pressure steam was used in the reboiler section. It is anticipated that this steam would be available at around 4.8 bar (APGT, 2015). An electrical load was considered for pumping. NaOH solution is required for the caustic treatment; for the Australian case, it was assumed that any pelletised NaOH used to prepare this solution would be imported. A MEA price of US$1.8/kg and a NaOH pellet price of US$385/tonne were used (CSIRO in-house chemical database). The price of steam, cooling water and electricity were taken from (Hosseini, Haque, Selomulya, & Zhang, 2016).

Cooling water is required to operate the crystalliser at a constant temperature. Electrical loads are also needed to run the hydro cyclones, centrifuges, dryers and pumps. Compressed air heating used direct contact electrical heaters. The solid KOH was assumed to be imported in a similar manner as the NaOH at a price of US$1200/tonne.

All capital and O&M costs in this paper are expressed in June 2015 Australian dollars similar to those considered in the reference reports (exchange rate of 1 AUD = 0.71 USD).

3. Cost comparison: CS-Cap process with conventional PCC technologies retrofitted to an Australian coal-fired power plant

The overall cost of the CS-Cap process is estimated, then compared to a conventional PCC + FGD configuration. As Australian coal-fired power plants do not have FGD units installed, our estimated CO2 capture cost includes the cost of FGD installation, whether for a new build or retrofitted power plant. The CS-Cap process cost is benchmarked against those reported previously for Australian coal-fired power plants using PCC + FGD. Reports prepared by Gamma Energy Technology and CO2CRC (CO2CRC, 2017) and Electric Power Research Institute (APGT, 2015) were used for cost comparison.

3.1. Assumptions, economic inputs and basis of reference reports for cost comparison

The retrofitted PCC facility evaluated in the (CO2CRC, 2017) study included two absorber trains, one regenerator per absorber train, one compression train per regenerator and eight reboilers per regenerator with CO_2 capture rate set at 90% with 24/7 operation. The PCC plant retrofit obtains steam for solvent regeneration via the crossover between intermediate-pressure and low-pressure systems in the existing plant. A backpressure steam turbine is introduced to step down the steam to the correct conditions for solvent stripping in the reboiler. The heat from the hot condensate returning from the reboiler supplements feedwater heating via heat exchangers. This base case PCC coal retrofit was a fully integrated MEA solvent facility. For the retrofit case, the existing pulverised-coal base plant without capture is a fully paid-off asset, in good condition with a suitably long life

(CO2CRC, 2017). The retrofit case assumes wet cooling and no upgrades are considered to the base plant.

Another report (APGT, 2015) details the cost and performance of black and brown-coal supercritical and ultra-supercritical coal-fired power plants with and without CCS. It includes the costs of FGD installation and of retrofitting PCC without FGD in an existing Australian black-coal-fired power plant. This information is used here to compare the CS-Cap case with other reference cases using a standard CO_2 capture process.

3.2. Scenarios comparing the CS-Cap process with conventional PCC + FGD installations

The following scenarios are detailed in SI:

- Scenario 1: Base case retrofitted with PCC, FGD + selective catalytic reduction (SCR)
- Scenario 2: Base case (with FGD & SCR and improved solvent)
- Scenario 3: Base case (with minimal FGD & SCR)
- Scenario 4: CS-Cap case with thermal reclamation
- Scenario 5: CS-Cap case with reactive crystallisation

3.3. Calculation basis for Total Plant Costs (TPC) of scenarios 1–5

The TPC in A$/kW for scenarios 1–3 with wet cooling (as applicable to black and brown coal) are taken from the (CO2CRC, 2017) report. Any PCC deployment cost in Australia has two components: CO_2 capture (PCC cost) and FGD cost for SO_2 capture. The cost breakup of TPC for PCC Retrofit + FGD/SCR with dry cooling is provided in (APGT, 2015). The TPC of PCC equipment retrofit + FGD/SCR for a wet cooling case is available (CO2CRC, 2017), however, no breakdown is available for the individual PCC retrofit and FGD/SCR costs.

To evaluate the PCC retrofit TPC in scenarios 4–5, the ratio of FGD/SCR cost requirements to the total TPC using dry cooling for a black-coal power plant is used to calculate the individual PCC equipment retrofit cost. An air-cooled condensing plant design can reduce plant gross output by as much as 1% compared with wet cooling, however, a similar cost ratio of 44–45% for FGD/SCR (dry cooling) is used here. The cost obtained for FGD/SCR was then subtracted from the combined TPC of the Scenario 1 PCC retrofit + FGD/SCR. The cost of the two CS-Cap regeneration techniques is then added to this modified TPC. This replaces the cost of FGD/SCR with the cost of the CS-Cap process.

3.4. Calculation basis for fixed and variable O&M costs for scenarios 4–5

The total fixed and variable costs in A$/kW-yr and A$/MWh for scenarios 1–3 have been taken from the report (CO2CRC, 2017). The individual fixed cost breakdown for PCC retrofit and FGD/SCR, used a similar factor of 44–45% as identified in (APGT, 2015). This is based on the cost estimation method, which assumes that the fixed part of the operating cost is taken as a percentage of the capital cost. By excluding FGD, the resulting 45% reduction in capital cost afforded the same reduction in fixed operating cost. The fixed O&M cost of regeneration techniques was then added to the fixed O&M cost of the PCC retrofit obtained from the report.

In the (APGT, 2015), FGD/SCR was found to provide 20% of the variable O&M costs. This factor was applied to the variable O&M costs obtained from the (CO2CRC, 2017). This allowed the FGD/SCR portion of the variable O&M costs to be replaced by those calculated for the CS-Cap process.

The TPC, operating and maintenance costs as outlined in SI Tables A, B, E and F have been calculated for a 900-MW gross power plant. However, the costs provided in (APGT, 2015) for the black-coal cases are calculated based on a 450-MW gross power plant with a net output of 331 MW with PCC. Hence, all costs mentioned in Tables A, B, E and F are converted to a similar basis for a 450-MW power plant, as shown in Table I (SI), to provide a similar basis for all scenarios to be compared. A power law equation commonly used to scale plant and equipment costs with changes in throughput or capacity has been used to adjust the costs (Haque & Somerville, 2013):

$$\text{New plant cost} = \text{Base Cost X (New capacity/base capacity)}^{\,0.65} \quad (1)$$

3.5. Calculation basis for LCOE & cost of CO_2 avoided

The LCOE is the average cost of producing electricity from any technology over its entire life. It is calculated by converting the capital and O&M costs of a plant into a cost per MWh.

The cost of electricity consists of four components: capital costs, O&M costs, fuel costs & CO_2 transportation and sequestration. When these costs are calculated independently, individual cost basis comparison can be misleading. To compare different size units, they must all have the same cost unit basis when combined to calculate the cost of electricity (typically $/MWh).

The criteria used to convert all costs including the capital, O&M costs into $/MWh is described in Section 17.2.5 of (APGT, 2015). In addition, the Average LCOE and cost of CO_2 avoided has been calculated as follows:

$$\begin{aligned} Average\ LCOE = {} & Finance\ Charges + Fixed\ O\&M + Variable\ Cost + Fuel\ Cost \\ & + Cost\ of\ Transportation\ \&\ Storage \qquad (2) \end{aligned}$$

$$\frac{Cost}{\mathrm{MWh}} = \frac{Cost \text{ in AUD per kW} \times\ Sent\ Output(kW)}{Sent\ Output(kW) \times Operation\ Hours\ \times\ Capacity factor} \times 1000 \quad (3)$$

All the various costs fixed or variable need to converted in to Cost/MWh to evaluate average LCOE.
Operation hours considered = 8760 per year,
Capacity factor=0.85

$$Avoided\ cost = \frac{\{LCOE_{with\ removal} - LCOE_{reference}\}\$/MWh}{\{CO2\ emissions_{reference} - CO2\ emissions_{with\ removal}\}tonnes/MWh} \qquad (4)$$

4. Results and discussion

4.1. Cost estimation results of regeneration options 1 and 2

As the incoming sulfur content in flue gases affects the capital and operating cost of the reclaimers, regeneration options 1 and 2 have been evaluated for both high and low-sulfur-content gases. Table 2 summarises the total costs for each scenario.

Supplementary Tables A and B detail the capital expenditure and operating costs of thermal reclamation when applied to the CS-Cap process for 700 ppm while Tables C and D cover 200ppm sulfur content in the incoming flue gas.

Supplementary Tables E and F detail the capital expenditure and operating cost of reactive crystallisation when applied to the CS-Cap process for 700 ppm while Tables G and H cover 200ppm sulfur content in the incoming flue gas.

Table 2: Total cost of reclamation techniques at 200ppm and 700ppm incoming sulfur content in flue gas

Description	Flue gas Sulfur (ppm)	Total Cost in AUD ex taxes were applicable	
		Capital($M)	Operating($M/yr)
Option 1 -Thermal reclaimer for one train of 900 MW plant	700	4.25	71.73
	200	2.26	37.59
Option 2 - Reactive crystalliser for one train of 900 MW plant	700	7.26	59.89
	200	7.05	26.19

Comparing costs outlined in Table 2, it can be seen that crystallisation is a cheaper regeneration technique than thermal reclamation for the CS-Cap process for any level of sulfur in the flue gas. The crystallisation capital requirements are nearly 1.7 times those of thermal reclamation due to the multiple process operations involved (supplementary Tables E-H). However, this cost is offset by the comparatively lower MEA losses in crystallisation which are approximately 0.5-0.6 times to that of thermal reclamation.

Unlike reactive crystallisation, which will not remove non-ionic degradation products, thermal reclamation is capable of removing most impurities. Due to the slow build-up of degradation products (apart from sulfate), periodic thermal reclaiming is likely still required in addition to crystallisation to remove non-ionic impurities. This research focused only on the removal of sulfate; hence, the comparison is based only on the efficiencies of sulfate removal from the spent absorbent.

Cost comparisons are provided next for a black coal fired plant (Table 3), then brown coal (Table 4) and both are summarised on a LCOE basis (Figure 2).

4.2. Cost comparison results of scenarios 1–5 (black-coal case)

To retrofit PCC to existing Australian coal-fired power plant, the total plant cost for CO_2 capture will include the TPC of PCC + FGD/SCR. This is applicable for scenarios 1–3 of Section 3. However, for CS-Cap scenarios 4–5, the TPC will include PCC retrofit + TPC of reclaimer (either thermal reclaimer or reactive crystalliser).

4.2.1. PCC retrofit cost & FGD/SCR cost capital cost

The cost of the regeneration techniques outlined in options 1 and 2 and calculated in Section 4.1 is for one train of the PCC system installed at a 900-MW gross power station. The net output of the plant is 662 MWe for a black-coal case.

Further, for CS-Cap case scenarios 4 and 5, the basic PCC equipment retrofit cost is slightly higher than scenarios 1–3. This is because the CS-Cap absorber column has to be larger than a standard column as a result of the additional SO_2 absorption section, which requires an additional packing height of 4 m. The CO_2 capture section is not altered, as the increased flue gas flow rate through the blower (depending on its location) and cooling duty applied to the direct contact cooler are only slight. The water balance in the SO_2 capture loop can be maintained by operating the column close to 40 °C.

Table 3: Total plant costs for retrofitting post-combustion capture (PCC) in a 450-MW (331 MW net) black-coal power plant

Cost	**Scenario 1**	**Scenario 2**	**Scenario 4**	**Scenario 5**
	Base case (with FGD & SCR)	Base case (with FGD & SCR and improved solvent)	CS-Cap case with thermal reclaimer	CS-Cap Case with reactive crystalliser
PCC equipment retrofit cost ($/kWe)	2,283	2,118	2,285	2,285
FGD + SCR retrofit cost ($/kWe)	1,817	1,742	24	42
Total plant cost (A$/kW sent out)	4,100	3,860	2,309	2,327
Total plant cost (A$bn)	1.9	1.4	1.0	1.0

FGD = flue gas desulfurisation; SCR = selective catalytic reduction

Table 3 enables the TPC comparison of scenarios 1–5 for the black-coal case. Scenario 3, i.e. PCC with minimal FGD/SCR, was not considered in the reference report and is not included here. Table 3 reveals that the capital cost of retrofitting the CS-Cap process, using a thermal reclaimer or crystalliser in an existing plant, would be the cheapest option for an Australian black-coal-fired plant. Within the accuracy of these cost estimates, the total plant costs for scenario 4 and 5 were found to be the same at A $1bn. As mentioned previously, the crystalliser will not remove all potential impurities, and hence periodic thermal reclamation may be needed, which will slightly raise the cost of Scenario 5. On the other hand, Scenario 4 does not need any additional batch thermal reclaimer, as the continuous thermal reclaimer of the CS-Cap process will be able to remove degradation products other than sulfate. The TPC reduction from Scenario 1 to Scenario 4 is around 43%. Note that scenarios 4 and 5 do not include any SCR installations. This is because the NOx component of the flue gas from Australian coal power plants is mostly “inert” NO (~99% NO, balance NO2 and N2O) (Meuleman, et al. 2010). NO does not react with amines, and hence passes through without affecting the CO_2 capture system. The FGD/SCR retrofit cost reductions are considerable when comparing Scenario 1 to scenarios 4 and 5.

The capital cost comparison in Table 3 definitely favours the CS-Cap process for Australian black-coal-fired power plants. However, it is important to calculate the LCOE, which includes the O&M, fuel, CO_2 transportation and sequestration costs as listed in SI tables K and N. This will identify whether the electricity produced could be cheaper, against standard technologies, if the CS-Cap process is used.

4.2.2. LCOE & CO_2-avoided cost for scenarios 1–5

Table J (SI) reveals the capture costs for retrofitting PCC to a black-coal power plant for scenarios 1–5. The data for scenarios 1 and 2 is taken from the reference report (CO2CRC, 2017). The various cost data for scenarios 4 and 5 have been calculated based on the method described in sections 3.2–3.4 and the data calculated in SI Table I. These retrofit costs in SI Table J are used to calculate the LCOE for our black-coal case (SI Table K). The average LCOE data for black coal case are summarised in Figure 2. Figure 2 reveals that the cost of electricity is comparable between the base cases and retrofitting either CS-Cap process.

4.3. Cost comparison results for scenarios 1–5 of Section 3 (brown-coal case)

4.3.1. PCC retrofit cost and FGD/SCR cost

Scenarios 1–2 are based on a supercritical 2100-MW brown-coal power plant with a net output of 1283 MW when retrofitted with PCC. Hence, for LCOE cost comparison, the costs of amine regeneration were converted to the larger plant size.

Further, the capital and operating costs for the thermal reclaimer and reactive crystalliser are calculated in SI tables C, D, G and H (200ppm Sulfur).The cooled flue gas flow rates for the brown and black coal cases are anticipated to be comparable for a similar capacity power plant. However, the brown coal power plant efficiency would reduce. A net capacity of 550 MW is assumed here for a 900-MW gross output brown-coal power plant for scenarios 4 and 5.

Table L (SI) shows the TPC and O&M costs adjusted for reclamation techniques applied to a 2100 MW (1283 MW net) brown-coal power plant for scenarios 1–5. The capital cost comparison in Table 4 again favours using the CS-Cap process for Australian brown-coal-fired power plants. Within the accuracy of these cost estimates, the total plant costs for scenario 4 and 5 were again found to be the same at A $5.8bn. However, the LCOE was calculated further below to identify whether the electricity produced would be cheaper against standard techniques if the CS-Cap process is used in Australian brown-coal-fired power plants.

Table 4: Total plant costs for retrofitting post-combustion capture (PCC) in a 2100-MW (1283 MW net) brown-coal power plant

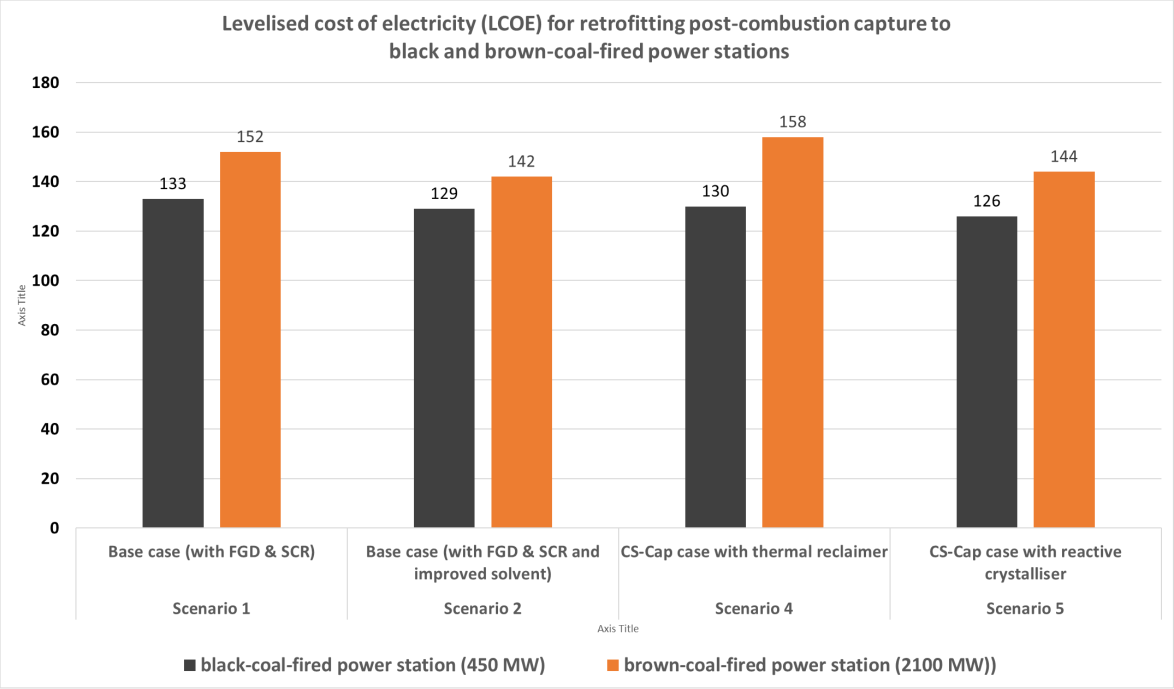
Cost	**Scenario 1**	**Scenario 2**	**Scenario 3**	**Scenario 4**	**Scenario 5**
	Base case (with FGD & SCR)	Base case (with FGD & SCR and improved solvent)	Base case (with minimal FGD & SCR), single boiler retrofit	CS-Cap case with thermal reclaimer	CS-Cap case with reactive crystalliser
PCC equipment retrofit cost ($/kWe)	2,728	2,579	2,728	2,730	2,730
FGD + SCR retrofit cost ($/kWe)	2,172	2,121	1172	9	29
Total plant cost (A$/kW sent out)	4,900	4,700	3,900	2,739	2,759
Total plant cost (A$bn)	10.3	9.9	1.5	5.8	5.8

FGD = flue gas desulfurisation; SCR = selective catalytic reduction

4.3.2. LCOE & CO_2-avoided costs for black and brown coals in scenarios 1–5

Table M (SI) shows the capture costs for retrofitting PCC to the brown-coal case for scenarios 1–5. The data for scenarios 1,2 and 3 is taken from the reference report (CO2CRC, 2017). The costs in (SI) Table M are used to calculate the LCOE for the brown-coal case (SI Table N) and are summarised here in Figure 2. For comparison, Figure 2 also includes the LCOE results for black coal scenarios discussed previously in section 4.2.2.

Figure2: - Levelised cost of electricity (LCOE) for retrofitting post-combustion capture to black and brown-coal-fired power stations



FGD = flue gas desulfurisation; SCR = selective catalytic reduction. For LCOE comparison, the calculated LCOE without CCS is $39/MWh with CO_2 emissions of 2.95 Mt/year for a Black coal 450-MW power plant. For a brown coal plant, the LCOE without CCS is $29/MWh with CO_2 emissions of 19.88 Mt/year for 2100-MW. Figure 2 reveals that the lowest cost of electricity is achieved by retrofitting the CS-Cap process using a reactive crystalliser to a 450-MW (331 MW net) black-coal power plant.

Further, in Figure 2, when comparing scenarios 1, 4 and 5, CS-Cap is also the cheapest option if used with a crystalliser in a brown-coal-fired power plant. The LCOE for scenario 2 and 3 is calculated only for single boiler retrofit (343 and 380 MW respectively) whereas scenario 1, 4 and 5 are calculated for full capacity (1283 MW) retrofit. The cost of retrofitting FGD+PCC reduces further if only a single boiler is retrofitted, as shown in scenarios 2 and 3. However, comparing CS-Cap cases for scenarios 4 and 5 against scenarios 2 and 3 is not valid. The CS-Cap cases are not evaluated here for single boiler retrofit, which will further reduce the cost of CS-Cap scenarios 4 and 5. The case of minimal FGD/SCR with a single boiler retrofit (Scenario 3) results in the lowest LCOE for the brown-coal case, as mentioned in the report (CO2CRC, 2017). This optimistic approach of applying minimal FGD/SCR as reported in(CO2CRC, 2017) is the cheapest option for Australian coal plants, but this theoretical outcome was not based on any bench or pilot-scale study. In contrast, the present CS-Cap costing study is based on the results of pilot and bench-scale testing and is considered more reliable. In addition, the potential cost benefit from selling by-products has not been included in this costing study. It might be possible to improve the economics of CS-Cap by selling K_2SO_4 generated as a by-product of the process.

Estimations of the overall CO_2 avoidance cost for both coal types are presented in Table 5

Table 5: Summary Table –CO_2 avoided cost for retrofitting post-combustion capture to black and brown-coal-fired power stations

Cost CO_2 avoided ($/tonne)	**Scenario 1**	**Scenario 2**	**Scenario 4**	**Scenario 5**
	Base case (with FGD & SCR)	Base case (with FGD & SCR and improved solvent)	CS-Cap case with thermal reclaimer	CS-Cap case with reactive crystalliser
black-coal-fired power station (450 MW)	**125**	**117**	**121**	**116**
brown-coal-fired power station (2100 MW)	**120**	**104**	**126**	**112**

Shifting the focus to modelling the cost of flue gas CO_2 removal instead of electricity reveals (Table 5) that only the regeneration using recrystallization (Scenario 5) produced lower values than the base case. CS-Cap incorporating reactive crystallisation is worthy of further investigation given its calculations are based on using the higher energy demand MEA solvent.

5. Conclusions and future work

High level modelling estimated the costs for the conceptual CS-Cap PCC process, with a focus on two alternative solvent recovery technologies. Both of the technologies evaluated experimentally in this paper can regenerate the sulfate-rich aqueous amine absorbent produced by the CS-Cap process. Due to its lower amine makeup requirements, modelling CS-Cap with reactive crystallisation revealed consistently lower LCOE and CO2 cost avoided values compared to thermal reclamation.

We found minimal differences in LCOE for a black-coal power plant comparing base case PCC ($133/MWh) with FGD and SCR against CS-Cap with a reactive crystalliser ($126/MWh) for both new build and retrofitting to an existing (450 MW) power plant. We can conclude that the CS-Cap process is unlikely to offer significant savings to Australian black-coal-fired power plants. The key factor being their higher SO_2 emissions compared to brown coal plants.

For a brown-coal 2100 MW power plant, the calculated LCOE is slightly lower for CS-Cap process using a crystalliser ($144/MWh) compared with a standard PCC retrofit process with FGD ($152/MWh).

The greatest contributor to the LCOE of CS-Cap is operating cost, rather than capital cost, due to its continuous and relatively high chemical requirements such as amine solvent makeup and alkali (KOH or NaOH) addition. The operating cost of regeneration increases by nearly 50% if sulfur levels rise from 200 to 700 ppm in the flue gases, e.g. resulting in an additional MEA loss of approx. 14% and an increase of 17-18% in CO_2 avoidance cost. For the lower sulfur brown coal, amine losses of ~15% for thermal reclamation and ~8–9% for reactive crystallisation are a significant operational cost.

Future work should therefore focus on further improving data quality and reducing operating costs by conducting experiments at pilot scale on brown coal plant.

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Declaration of interests

☐ The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

☐The authors declare the following financial interests/personal relationships which may be considered as potential competing interests:

CRediT author statement

Bharti Garg, Data curation, Writing- Original draft preparation

Nawshad Haque, Methodology, Software

Ashleigh Cousins, Supervision, Reviewing

Pauline Pearson, Reviewing and Editing

T Vincent Verheyen, Supervision, Writing- Reviewing and Editing,

Paul HM Feron, Visualization, Conceptualization

Supplementary Information

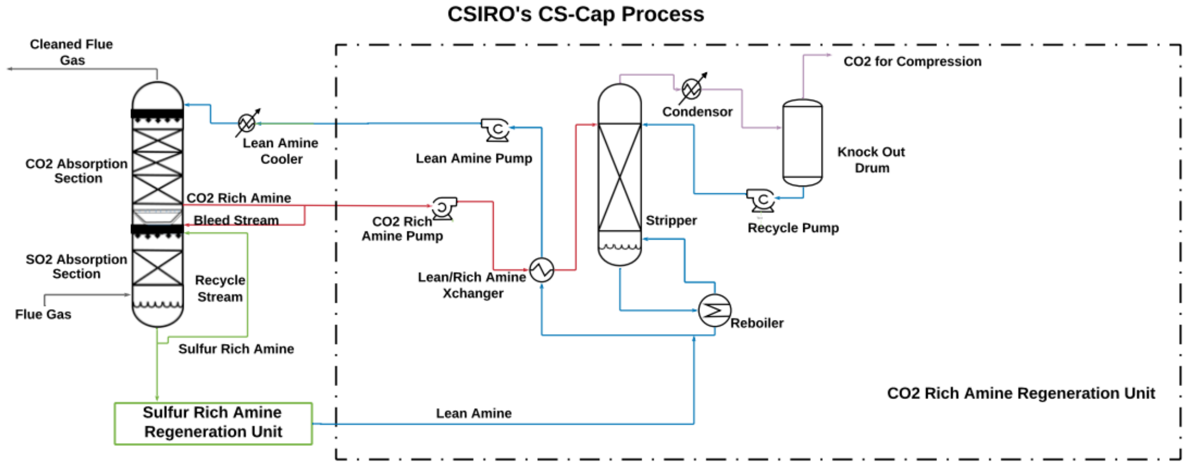

Figure A: Process flow diagram of CSIRO’s CS-Cap process

Scenarios comparing the CS-Cap process with conventional PCC + FGD installations

Scenario 1: Base plant retrofitted with PCC, FGD + selective catalytic reduction (SCR)

The design basis for the base case in (Retrofitting CCS To Coal : Enhancing Australia's Energy Security Results And Discussions, 2017) is from (Australian Power Generation Technology Report, 2015), and consists of a pulverised-coal case with 85% capacity factor. Key pieces of equipment included in the base power plant were pulverised-coal boilers, combustion-turbine generators, steam-turbine generator, circulating water pumps and drivers, cooling systems, condensers, air-separation units and main transformers.

The sites chosen for PCC retrofit studies were a generic greenfield site in the Latrobe Valley in Victoria, Australia for the brown-coal case, and a New South Wales location for the black-coal case at an elevation of 110 m. No nearby railway line or road was considered in the study for fuel delivery, as the sites were assumed to be at mine mouth. Dry-cooling and ash-handling systems were considered, due to a shortage of supply for cooling water in some parts of Australia.
The CO_2 transportation pipeline and storage sites were not included in the capital cost estimates; however, a $15/tonne CO_2 cost was added to the LCOE for transportation and storage.

The TPC included the cost of equipment (complete with initial chemical and catalyst loadings), materials, labour (direct and indirect), engineering and construction management, and contingencies (process and project). The owner's costs were excluded.

The O&M costs had fixed and variable components. The fixed components included operating labour, maintenance material and labour, and administrative and support labour. The variable components included consumables, waste disposal, co-product or by-product credit and fuel costs. For the purposes of the O&M cost analysis, the value of these waste streams was assumed to be zero.
The project and process contingency, other exclusions from the cost estimate, maintenance, material, and other cost details can be found in Chapter 15 of (Australian Power Generation Technology Report, 2015).

The new PCC plant in the base case Scenario 1 included the cost of FGD and SCR, as these units are missing from Australian coal-fired power plants.

Scenario 2: Base case (with FGD & SCR and improved solvent)

This scenario had a similar design basis as Scenario 1, but used an improved solvent for CO_2 absorption. The improved solvent was assumed to require less steam to regenerate, and hence will raise the net MWe output of the power plant. This case was considered based on the confidence gained from learning by doing and reduced capital requirements resulting from the performance of advanced solvents in several CCS projects worldwide.

Scenario 3: Base case (with minimal FGD & SCR)

The scenario was similar to Scenario 1, but did not include the cost of deep FGD & SCR. This scenario was particularly for Victorian brown coal with low sulfur content. The use of minimal FGD and SCR is thought to significantly reduce the capital requirements, resulting in a lower LCOE. Minimal FGD and SCR may include a direct contact cooler, along with alkali treatment for SO_2 removal (Retrofitting An Australian Brown Coal Power Station With Post-Combustion Capture, 2018) and installing low NOx burners instead of SCR. This reduces the cost of adding PCC by approximately 45% for a black-coal-fired power plant (Australian Power Generation Technology Report, 2015). Hence, a similar assumption has been considered for a brown-coal case. The retrofitting report, however, does not provide the cost for Scenario 3 for black coal; hence, this case has not been considered for black coal here.

Scenario 4: CS-Cap case with thermal reclamation

This scenario included PCC retrofit in a base plant with no FGD or SCR. The FGD was replaced by the CS-Cap process combined with thermal reclamation as discussed in Section 2.1.1. The SCR was not included because of the very low NO_2 concentrations in the total NOx typically emitted (~99% NO, balance NO_2 and N_2O) in Victorian coal-fired flue gases. NO does not react with amines, and hence passes through without affecting the CO_2 capture system.

Scenario 5: CS-Cap case with reactive crystallisation

This scenario included PCC retrofit in a base plant with no FGD or SCR. The FGD was replaced by the CS-Cap process combined with reactive crystallisation as discussed in Section 2.1.2. SCR costs were not included for the Victorian coal case.

Table A: Capital cost of thermal reclaimer for one train of 900 MW gross power station (700 ppm sulfur)

Direct plant costs	**Basis (DEC = EPC + freight)**	**$M ex GST**	**Assumptions**
Equipment purchase	EPC	0.92	From APEA (AUD)
Freight	% of (EPC)	0.09	10%
DEC		1.01	
Installation	% of (DEC)	0.46	45%
Instrumentation	% of (DEC)	0.25	25%
Minor piping	% of (EPC)	0.15	16%
Structural	% of (EPC)	0.14	15%
Electrical	% of (DEC)	0.25	25%
Buildings	% of (EPC)	0.23	25%
Yard improvements	% of (EPC)	0.14	15%
Service facilities	% of (EPC)	0.37	40%
HSE functions	% of (EPC)	0.09	10%
Subtotal		3.09	
Total indirect costs			
Land			
Engineering supervision	% of (DEC)	0.51	50%
Legal expenses	% of (DEC)	0.04	4%
Construction expenses	% of (DEC)	0.41	40%
Subtotal so far		4.05	
Working capital			
Working capital		0.20	5%
Fixed capital investment		4.25	
Total capital (ex GST)		**4.25**	

The equipment cost has been taken from Aspen Plus Economic Analyser (APEA) in USD and converted into AUD at an exchange rate of 1AUD= 0.71USD

Table B: Operating cost of thermal reclaimer for one train of 900 MW gross power station (700 ppm sulfur)

Operating costs summary				
	Item	**Total cost ($M)**	**Consumption t/yr**	**Price per unit (AUD)**
Raw materials	MEA	13.79	5,441	$2.5/kg
	NaOH	52.328	96,500	$542/t
Utilities	Electricity	0.0001	514 MWh/yr	$0.1/MWh
	Steam	4.41	127,475 MWh/yr	$34.57/MWh
	Cooling water	0.00	0	$2.5/cubic metre
	Air	0.00	0	Free
	Item		**Assumptions**	**Price per unit**
Total fixed charges	Labour	0.39	2 people for operation 1 for maintenance	$73/h
	Maintenance and repairs	0.21	5% of total capital cost	NA
	Operating supplies	0.04	1% of total capital cost	NA
	Taxes (property)	0.08	2% of total capital cost	NA
	Insurance	0.04	1% of total capital cost	NA
Depreciation & capital	Fixed capital depreciation	0.21	5% of total capital cost	NA
	Interest on capital	0.21	5% of total capital cost	NA
Total product cost		**71.73**		

The fresh MEA requirement is calculated based on the simulation result for the worst case of 5 M MEA with 16–17 wt% sulfate in the recycled stream sent for regeneration. Based on experimental and modelling results, the worst-case MEA losses determined are 15–16 wt%. The amine required to compensate for this loss has been added to the chemical requirements.

Table C: Capital cost of thermal reclaimer for one train of 900 MW gross power station (200 ppm sulfur)

Direct plant costs	**Basis (DEC = EPC + freight)**	**$M ex GST**	**Assumptions**
Equipment purchase	EPC	0.49	From APEA (AUD)
Freight	% of (EPC)	0.05	10%
DEC		0.54	
Installation	% of (DEC)	0.24	45%
Instrumentation	% of (DEC)	0.14	25%
Minor piping	% of (EPC)	0.08	16%
Structural	% of (EPC)	0.07	15%
Electrical	% of (DEC)	0.14	25%
Buildings	% of (EPC)	0.12	25%
Yard improvements	% of (EPC)	0.07	15%
Service facilities	% of (EPC)	0.20	40%
HSE functions	% of (EPC)	0.05	10%
Subtotal		1.65	
Total indirect costs			
Land			
Engineering supervision	% of (DEC)	0.27	50%
Legal expenses	% of (DEC)	0.02	4%
Construction expenses	% of (DEC)	0.22	40%
Subtotal so far		2.15	
Working capital			
Working capital		0.11	5%
Fixed capital investment		2.26	
Total capital (ex GST)		**2.26**	

The equipment cost has been taken from Aspen Plus Economic Analyser (APEA) in USD and converted into AUD at an exchange rate of 1AUD= 0.71USD

Table D: Operating cost of thermal reclaimer for one train of 900 MW gross power station (200 ppm sulfur)

Operating costs summary				
	Item	**Total cost ($M)**	**Consumption t/yr**	**Price per unit(AUD)**
Raw materials	MEA	12.17	4,801	$2.5/kg
	NaOH	20.188	37,230	$542/t
Utilities	Electricity	0.000048	477 MWh/yr	$0.1/MWh
	Steam	4.41	127,475 MWh/yr	$34.57/MWh
	Cooling water	0.00	0	$2.5/cubic metre
	Air	0.00	0	Free
	Item		**Assumptions**	**Price per unit**
Total fixed charges	Labour	0.39	2 people for operation 1 for maintenance	$73/h
	Maintenance and repairs	0.11	5% of total capital cost	NA
	Operating supplies	0.02	1% of total capital cost	NA
	Taxes (property)	0.05	2% of total capital cost	NA
	Insurance	0.02	1% of total capital cost	NA
Depreciation & capital	Fixed Capital Depreciation	0.11	5% of total capital cost	NA
	Interest on capital	0.11	5% of total capital cost	NA
Total product cost		**37.59**		

Table E: Capital cost of reactive crystalliser for one train of 900 MW gross power station (700 ppm sulfur)

Capital expenditure			
Direct plant costs	**Basis (DEC = EPC + freight)**	**$M ex GST**	**Assumptions**
Equipment purchase	EPC	1.57	From APEA (AUD)
Freight	% of (EPC)	0.16	10%
DEC		1.73	
Installation	% of (DEC)	0.78	45%
Instrumentation	% of (DEC)	0.43	25%
Minor piping	% of (EPC)	0.25	16%
Structural	% of (EPC)	0.24	15%
Electrical	% of (DEC)	0.43	25%
Buildings	% of (EPC)	0.39	25%
Yard improvements	% of (EPC)	0.24	15%
Service facilities	% of (EPC)	0.63	40%
HSE functions	% of (EPC)	0.16	10%
Subtotal		5.28	
Total indirect costs			
Land			
Engineering supervision	% of (DEC)	0.87	50%
Legal expenses	% of (DEC)	0.07	4%
Construction expenses	% of (DEC)	0.69	40%
Subtotal so far		6.91	
Working capital			
Working capital		0.35	5%
Fixed capital investment		7.26	
Total capital (ex GST)		**7.26**	

The equipment cost has been taken from Aspen Plus Economic Analyser (APEA) in USD and converted into AUD at an exchange rate of 1AUD= 0.71USD

Table F: Operating cost of reactive crystalliser for one train of 900 MW gross power station (700 ppm sulfur)

Operating costs summary				
Variable cost	**Item**	**Total cost ($M/yr)**	**Amount used**	**Price per unit (AUD)**
Raw materials	MEA	9.20	3,628 t/yr	$2.5/kg
	KOH	48.024	64,333 t/yr	$530/t
Utilities	Electricity	0.00	1,816 MWh/yr	$0.1/MWh
	Steam	0.00	0	$34.57/MWh
	Cooling water	0.76	303,683 m^3/yr	$2.5/cubic metre
	Air	0.00	0	Free
Total variable O& M		57.98		
	Item	**$M/year**	**Assumptions**	**Price per unit**
Total fixed charges	Labour	0.53	3 people for operation 1 for maintenance	$73/h
	Maintenance and repairs	0.36	5% of total capital cost	NA
	Operating supplies	0.07	1% of total capital cost	NA
	Taxes (property)	0.15	2% of total capital cost	NA
	Insurance	0.07	1% of total capital cost	NA
Depreciation & capital	Fixed capital depreciation	0.36	5% of total capital cost	NA
	Interest on capital	0.36	5% of total capital cost	NA
Fixed O&M		1.18		
Total product cost		**59.89**		

The fresh MEA requirement was calculated based on MEA losses during the separation of the mother liquor from the crystals during filtration. There are some losses during drying as well. As per the simulation, MEA losses are insignificant, but in actual experimental results, the total loss of mother liquor is around 8–10%. This results ~1–2% loss of MEA, depending upon the concentration of MEA in the spent absorbent sent for regeneration. At large scale, the industrial filters and centrifuges are unlikely to achieve 100% separation of crystals from the mother liquor. Hence, the actual loss of MEA as determined in experimental work has been used as a worst-case scenario for the MEA makeup requirements here.

Table G: Capital cost of reactive crystalliser for one train of 900 MW gross power station (200 ppm sulfur)

Capital expenditure			
Direct plant costs	**Basis (DEC = EPC + freight)**	**$M ex GST**	**Assumptions**
Equipment purchase	EPC	1.53	From APEA (AUD)
Freight	% of (EPC)	0.15	10%
DEC		1.68	
Installation	% of (DEC)	0.76	45%
Instrumentation	% of (DEC)	0.42	25%
Minor piping	% of (EPC)	0.24	16%
Structural	% of (EPC)	0.23	15%
Electrical	% of (DEC)	0.42	25%
Buildings	% of (EPC)	0.38	25%
Yard improvements	% of (EPC)	0.23	15%
Service facilities	% of (EPC)	0.61	40%
HSE functions	% of (EPC)	0.15	10%
Subtotal		5.13	
Total indirect costs			
Land			
Engineering supervision	% of (DEC)	0.84	50%
legal expenses	% of (DEC)	0.07	4%
Construction expenses	% of (DEC)	0.67	40%
Subtotal so far		6.71	
Working capital			
Working capital		0.34	5%
Fixed capital investment		7.05	
Total capital (ex GST)		**7.05**	

The equipment cost has been taken from Aspen Plus Economic Analyser (APEA) in USD and converted into AUD at an exchange rate of 1AUD= 0.71USD

Table H: Operating cost of reactive crystalliser for one train of 900 MW gross power station (200 ppm sulfur)

Operating costs summary				
Variable cost	**Item**	**Total cost ($M/yr)**	**Amount used**	**Price per unit (AUD)**
Raw materials	MEA	8.02	3,163 t/yr	$2.5/kg
	KOH	16.01	21,444 t/yr	$530/t
Utilities	Electricity	0.00012	1,191 MWh/yr	$0.1/MWh
	Steam	0.00	0	$34.57/MWh
	Cooling water	0.29	117,430 m^3/yr	$2.5/cubic metre
	Air	0.00	0	Free
Total variable O&M		24.32		
	Item	**$M/year**	**Assumptions**	**Price per unit**
Total fixed charges	Labour	0.53	3 people for operation 1 for maintenance	$ 73/hr
	Maintenance and repairs	0.35	5% of total capital cost	NA
	Operating supplies	0.07	1% of total capital cost	NA
	Taxes (property)	0.14	2% of total capital cost	NA
	Insurance	0.07	1% of total capital cost	NA
Depreciation & capital	Fixed capital depreciation	0.35	5% of total capital cost	NA
	Interest on capital	0.35	5% of total capital cost	NA
Fixed O&M		1.16		
Total product cost		**26.19**		

Table I: Total plant and O&M costs adjusted for reclamation techniques applied to a 450-MW (331 MW net) black-coal power plant (700 ppm sulfur content)

Capital cost		**Thermal reclamation**	**Reactive crystallisation**
One train TPC for 662 MW sent out	A$	4,247,562.9	7,256,037.0
TPC for three trains	A$	12,742,688.8	21,768,110.9
Factor of cost reduction from 662 MW to 331 MW		0.6	0.6
TPC for three trains for 331 MW sent out	A$	8,120,664.7	13,872,388.6
TPC	A$/kW	24.5	41.9
Operating costs			
Fixed			
Fixed cost for one train	A$	776,480.7	1,178,643.3
Fixed cost for three trains	A$	2,329,442.0	3,535,930.0
Fixed cost for three trains	A$/kW-yr	3.5	5.3
Factor of cost reduction from 662 MW to 331 MW		0.6	0.6
Fixed cost for three trains for 331 MW sent out	AS/kW-yr	2.2	3.4
Variable			
Cost for one train	A$/MWh	14.3	12.1
Cost for three trains	A$/MWh	42.9	36.4
Factor of cost reduction from 662 MW to 331 MW		0.6	0.6
Cost for three trains for 331 MW sent out	A$/MWh	27.4	23.2

Table J: Post-combustion capture retrofitting costs in a 450-MW (331 MW net) black-coal power plant

Cost	**Scenario 1**	**Scenario 2**	**Scenario 4**	**Scenario 5**
	Base case (with FGD & SCR)	Base case (with FGD & SCR and improved solvent)	CS-Cap case with thermal reclaimer	CS-Cap case with reactive crystalliser
Total plant cost (A$/kW sent out)	4,100	3,860	2,307	2,325
Fixed operation & maintenance (A$/kW-year)	75	70	44	45
Variable (A$/MWh)	11	11	36	32

Table K: Levelised cost of electricity (LCOE) for retrofitting post-combustion capture to a black-coal-fired power station (450 MW)

Cost	**Scenario 1**	**Scenario 2**	**Scenario 4**	**Scenario 4**
	Base case (with FGD & SCR)	Base case (with FGD & SCR and improved solvent)	CS-Cap case with thermal reclaimer	CS-Cap case with reactive crystalliser
Finance charges	55	52	31	31
Fixed operation and maintenance	10	9	6	6
Variable cost	11	11	36	32
Fuel cost	41	41	41	41
Cost of T&S	16	16	16	16
Cost of carbon	0	0	0	0
Average LCOE ($/MWh)	**133**	**129**	**130**	**126**
CO_2 emitted (Mt/year)	0.32	0.3	0.32	0.32
MWnet	331	356	331	331
CO_2 emissions (tonnes/MWh)	0.13	0.11	0.13	0.13
CO_2 captured ($/tonnes)	104	97	100	95
CO_2 avoided ($/tonnes)	**125**	**117**	**121**	**116**

FGD = flue gas desulfurisation; SCR = selective catalytic reduction. For comparison, the LCOE without CCS is $39/MWh with CO_2 emissions of 2.95 Mt/year for a 450-MW power plant.

Table L: Total plant and O&M costs adjusted for reclamation techniques applied to a 2100-MW (1283 MW net) brown-coal power plant (200 ppm sulfur content)

Capital cost		**Thermal reclamation**	**Reactive crystallisation**
One train TPC for 550 MW sent out	A$	2,262,684.0	7,047,683.1
TPC for three trains	A$	6,788,052.1	21,143,049.3
Factor of cost increase from 550 MW to 1283 MW		1.7	1.7
TPC for three trains for 1283 MW sent out	A$	11,772,176.0	36,667,322.8
TPC	A$/kW	9.2	28.6
Operating costs			
Fixed			
Fixed cost for one train	A$	597,841.6	1,159,891.5
Fixed cost for three trains	A$	1,793,524.7	3,479,674.4
Fixed cost for three trains	A$/kW-yr	3.3	6.3
Factor of cost increase from 550 MW to 1283 MW		1.7	1.7
Fixed cost for three trains for 1283 MW sent out	AS/kW-yr	5.7	11.0
Variable			
Cost for one train	A$/MWh	9.0	6.1
Cost for three trains	A$/MWh	26.9	18.3
Factor of cost increase from 550 MW to 1,283 MW		1.7	1.7
Cost for three trains for 1,283 MW sent out	A$/MWh	46.7	31.8

Table M: Post-combustion capture retrofitting costs in a 2100-MW (1283 MW) brown-coal power plant

Cost	**Scenario 1**	**Scenario 2**	**Scenario 3**	**Scenario 4**	**Scenario 5**
	Base case (with FGD & SCR)	Base case (with FGD & SCR and improved solvent)	Base case (with minimal FGD & SCR)	CS-Cap case with thermal reclaimer	CS-Cap case with reactive crystalliser
Total plant cost (A$/kW sent out)	4,900	4,700	3,900	2,738	2,757
Fixed operation and maintenance (A$/kW-year)	70	65	60	45	50
Variable (A$/MWh)	14	14	12	58	43

Table N: Levelised cost of electricity (LCOE) for retrofitting post-combustion capture to brown-coal-fired power station (2100 MW)

Cost	**Scenario 1**	**Scenario 2**	**Scenario 3**	**Scenario 4**	**Scenario 5**
	Base case (with FGD & SCR)	Base case (with FGD & SCR and improved solvent)	Base case (with minimal FGD & SCR)	CS-Cap case with thermal reclaimer	CS-Cap case with reactive crystalliser
Finance charges	63	60	50	35.19	35.43
Fixed operation and maintenance	9	9	8	5.9	6.5
Variable cost	14	14	12	58	43
Fuel cost	38	34	31	34	34
Cost of T&S	28	25	23	25	25
Cost of carbon	0	0	0	0	0
Average LCOE($/MWh)	**152**	**142**	**124**	**158**	**144**
CO_2 emitted (Mt/year)	2.34	0.47	0.47	2.34	2.34
MWe	1,283	343	380	1283	1283
CO_2 emissions (tonnes/MWh)	0.24	0.18	0.17	0.24	0.24

CO_2captured($/tonne)	93	81	65	101	88
CO_2 avoided ($/tonne)	**120**	**104**	**86**	**126**	**112**

The LCOE without CCS has been used as $29/MWh with CO_2 emissions of 19.88 Mt/year for a 2100-MW power plant.